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RESEARCH STUDY OF COAL PREPARATION PLANT AND BY-PRODUCT COKE PLANT EFFLUENTS

CF AND I STEEL CORPORATION

PREPARED FOR
ROBERT S. KERR ENVIRONMENTAL
RESEARCH LABORATORY

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RESEARCH STUDY OF COAL PREPARATION PLANT AND BY-PRODUCT COKE PLANT EFFLUENTS

by

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Project 12010 DNF
Program Element 1BB036

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Source stream characterizations and bench/pilot treatability studies were performed on waste water from an integrated coal washery/by-product coking operation at an iron and steel plant. Major parameters of interest were phenolics, cyanide, ammonia, and suspended solids. The feasible treatment schemes identified among nine procedures tested were: biological activated sludge and carbon adsorption (for phenolics and cyanide); steam stripping/acid adsorption (for ammonia); and neutralization/polyelectrolyte addition/centrifugation (for coal suspensions). In addition, a new full-scale design for the coal preparation plant was produced.

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ABSTRACT

Studies were conducted to determine feasible methods for treatment of waste effluents from a Coal Preparation and By-Product Coke Plant.

A survey of the effluent streams pinpointed water pollution sources, degrees and types of contaminants, and waste stream volumes.

Laboratory and pilot studies using plug flow and complete mix activated sludge systems for treating the combined effluent proved to be feasible. Reasonably consistent phenol reduction to less than 1 ppm and cyanide reduction to 1-5 ppm were obtained. Difficulties were experienced in obtaining a sludge with good settling characteristics, probably due to variations in phenol concentrations and the presence of toxic substances in the waste stream.

Another feasible treatment method determined was adsorption using activated carbon. Standard carbon columns were used to remove phenol, color and other organics followed by a newly developed catalyst, oxygen, and activated carbon treatment method for cyanide destruction.

The only feasible treatment determined for ammonia removal was steam stripping of the concentrated ammonia flushing liquor stream, with return of the stripped ammonia to the by-product gas treatment process for removal by acid adsorption.

Other methods tested or investigated through a review of the literature were Biological Trickling Filter, Ion Exchange, Electrodialysis, Reverse Osmosis, Chemical destruction by Ozone, Chlorine, and Chlorine Dioxide.

The Coal Preparation Plant waste was primarily a solids removal and sludge disposal problem. Neutralization of the waste effluent with spent pickle acid, plus the addition of a polyelectrolyte, produced sufficient settling velocity for a clarifier-centrifuge system in a completely closed system layout. Design of full scale facilities was completed.

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CONCLUSIONS

1. This study was based on the conditions at the Pueblo Plant of CF&I Steel Corporation and while the facilities and operations of most By Product Coke Plants at integrated steel mills are similar, local variations do exist. Therefore the study, while believed to be generally applicable, may not entirely reflect the conditions and problems at any other specific plant.

2. The water effluent from the Coke Plant area is probably the most contaminated effluent from an integrated steel plant. In addition to suspended and settleable material, the effluent contains highly variable amounts of dissolved chemical contaminants. Principal of these are various ammonia, cyanide and phenolic compounds.

3. There are several waste water streams entering the existing Coke Plant sewer system which should be eliminated or diverted before final treatment of the total effluent water. These include both uncontaminated water needing no treatment and contaminated water which can best be treated or recycled at the source.

4. The coal washery pollution problem can be solved by modifications and additions to the coal washery and a closed water recirculation system as described in the report, provided the described operating procedures are also followed. In addition to solving the water pollution problem, additional fine coal will be recovered which will improve the cost benefit ratio of the project.

5. Several potentially satisfactory methods of treating the Coke Plant final effluent water for removal of phenol and cyanide were indicated by the various small bench and laboratory scale pilot treatment plants studied. These include various variations of biological treatment and treatment with activated carbon. The study indicates activated carbon treatment may be the best treatment method. However, further study must be made particularly in the area of operating and construction costs of the alternate treatment methods before full scale plant construction. In the case of the biological processes in particular, larger scale pilot plant studies indicative of actual operating conditions should be made on the representative final effluent stream after the recommended process and sewer system modifications have been completed. This further study will define the parameter of interferences and limitations indicated before installation of a full scale treatment plant.

6. A primary treatment for removal of settleable solids and floating materials should precede the final treatment facility. Retention time required may vary from one hour to one day dependent upon the final treatment method. The one hour detention time is sufficient for solids and oil removal, with a one day retention possibly being required to equalize contaminant concentration levels ahead of a biological treatment plant to avoid shock loading. The exact detention time required can not be determined until the process and sewer system changes are accomplished.

7. No potentially satisfactory method of ammonia removal in a waste treatment plant was found. The best method of ammonia removal from the plant effluent appears to be steam stripping of ammonia from the concentrated source streams in the normal by product gas treatment plant and returning the ammonia to the process for acid absorption in the existing saturators.

RECOMMENDATIONS

The coal washery treatment facilities described should be installed and operated as further defined in this report. The treatment facilities will eliminate the discharge of contaminants from the coal washery by utilizing chemical treatment, solids separation and sludge dewatering. The process water will be recirculated, with adequate blowdown provided by the amount of water held by the washed coal and solid waste. A significant amount of usable fine coal will also be recovered. Treatment of this water in a separate system is necessary because of the large quantities of suspended solids which are difficult to remove and because there are differences in settling characteristics of this waste and the chemically contaminated coke plant waters.

Other than the coal washery, there are several other effluent streams which should be isolated from the plant sewer system prior to final treatment in order to reduce the flow and the loading to a final treatment plant. There are several significant uncontaminated effluent streams which can be discharged without treatment by diversion to an existing clean water sewer. Minor piping changes would be required. In addition the study indicates that several contaminated streams should be treated separately. This is noted below and described elsewhere in the report.

By far the largest ammonia source in the coke plant effluent is the ammonia flushing liquor discharged by sewer. Ammonia in the excess ammonia flushing liquor should be released by steam stripping and returned to the process for removal by acid adsorbtion in the existing saturators. Existing ammonia stills can be refurbished to meet the stripping requirements. No other practical means of ammonia removal was determined in the study; however, other research projects are underway such as at Armco Steel Corporation's Houston Works which may lead to possible future improvement.

The spent caustic from the secondary light oil recovery process should be recycled. A recovery of caustic plus additional recovery of light oil is feasible.

It is recommended that the waste water effluent from the light oil separator be cooled by means of a cooling tower and returned to the final cooler to replace part of the mixture of uncontaminated primary cooler water and plant water now in use. Any light oil contained in this water would be recycled to the absorber following the final cooler and reduce the load to a final treatment plant.

Excess plant water is presently being used at the two coke quenchers. This water requirement should be met with either final cooler water or plant makeup water now in use. Any light oil contained in this water would be recycled to the absorber following the final cooler and reduce the load to a final treatment plant.

Excess plant water is presently being used at the two coke quenchers. This water requirement should be met with either final cooler water or plant makeup water, subject to the results of full scale testing to determine operational problems. With the exception of the coal washery improvements, the recommended process and piping changes are relatively straight forward and inexpensive. It is therefore recommended that these changes be accomplished during a later study which will define the operating and capital costs of a full scale treatment plant for the removal of phenol, cyanide and other contaminants. Studies will include the activated carbon and activated sludge methods, as determined to be potentially suitable for full scale treatment, and the design information developed in this study.

Upon completion of the proposed piping and process changes, further study should be instituted to refine the design parameters. Completion of the coal washery phase would not be necessary since this is an intermittent process and its discharge could be excluded from the investigations. The extent of this refinement would depend upon which treatment method were chosen as a result of the cost effectiveness study. Design of a full scale activated carbon plant would require very little investigative work. Design of an activated sludge plant would require a three to four month larger-scale pilot plant study to further determine and solve the operating and efficiency problems encountered and described in this report.

INTRODUCTION

Project Need

The CF&I Steel Corporation is pursuing a pollution abatement program at its Pueblo Plant so that the water discharged into the Arkansas River will result in stream quality conducive to domestic, industrial, and agricultural purposes. At the point of CF&I discharge, the Arkansas River with its very meager flow is presently classified for agricultural and industrial quality. A few miles down stream the river is classified for domestic quality as well.

CF&I has made continuing progress on its water quality program, has worked closely with the State officials and has received compliments several times from the Colorado Water Pollution Control Commission and its predecessor and co-agency - the Colorado Department of Public Health. Nevertheless, the effluents from the coke plant area present a major problem in the attainment of the overall objectives. Satisfactory treatment of coke plant effluents is an industry wide problem and one for which no satisfactory solutions have been demonstrated.

In order to accomplish the overall objective, it was necessary to develop techniques to treat coke plant effluent efficiently and economically. As a first step in this direction CF&I developed a project to characterize the wastes involved and to bench test several possible treatment methods.

The Pollution Control Department of Nalco Chemical Company, Chicago, Illinois was engaged as a consultant to provide technical advice and services in the conduct of the overall study. The Water Management Division of Calgon Corporation, Pittsburgh, Pennsylvania was engaged as a consultant on activated carbon treatment and performed some chemical analyses. Stearns-Roger Corporation, Denver, Colorado was engaged as a consultant on the coal washery studies and the engineering design for the full scale plant. CF&I engineers, chemists, and technicians performed most of the field work and chemical analyses required in the study. All of the aforementioned consultants provided detailed written reports to form the principal basis for this final report.

FACILITIES DESCRIPTION

Coal Washery

Colorado coals, which were laid down during the cretaceous age, are primarily of the Sub-Bituminous and Bituminous groups. As compared with the older Eastern coals, Colorado's coal is generally higher in ash and of poorer coking quality. In order to produce a satisfactory metallurgical coke two practices are followed which are common in the coking industry:

1. A small amount (usually between 10 and 15 percent) of low volatile coal is blended with CF&I's high volatile coals to improve the physical characteristics of the coke.

2. The raw coals as mined contain mineral impurities which increase the ash content of the coal; therefore the coal is washed to reduce the amount of these impurities, thereby improving both the physical and chemical characteristics of the coke.

The cleaning of coal (commonly termed washing) consists of separating the mine output into two products; coal sufficiently cleaned to be utilized directly, and the other, the high-ash minerals present in the coal. This operation performed at the Pueblo Plant rather than at the mine site, is necessary in modern mining practice to more economically remove the impurities in a washing plant than to attempt to do it underground during the loading process. At the Pueblo Plant of CF&I, the percentage of mine output unusable as fuel or in coke production is approximately 20-25%.

Impurities in coal may have resulted from a number of causes--residual inorganic constituents in the original vegetation, sand or dirt blown into the coal-forming mass, deposits of sediment forming partings, pyrite sulphur formed by interaction of the coal-forming mass and water containing ferrous sulphate, or saline deposits from water. Of the various forms of impurities the residual inorganics (inherent ash) are such an intimate part of the coal as to make their removal impossible. Ordinarily the remaining forms of impurities can be separated from the coal by rather simple methods. In the case of extremely fine material the separation may be made by flotation, in which the coal particles, because of different surface characteristics, are separated from the rock.

High volatile coal is now being received at the Pueblo Plant from the company-owned Allen Mine in the Trinidad Colorado area; low-volatile coal is purchased from non-CF&I sources. These coals are crushed to minus 2½" at the mines and proportioned before washing the mixture. The liberation of free impurity from laminated material is a minor problem.

A flow diagram of the CF&I Pueblo Plant coal washery is presented on the following page (Figure 1). The primary separation of the raw coal into washed coal and refuse takes place in two air-operated Jeffrey-Baum type jigs, each jig having three two-cell compartments. Waste material which settles in these jigs is removed by elevators and is transported by belt conveyors to a refuse storage bin. The washed coal is dewatered as it proceeds through a combination of wedge wire stationary screens and double-deck vibrating screens where plus 1/4 inch coal is delivered directly to the washed coal belt and 1/4 inch x 48 mesh coal is transported to basket type centrifugal dryers. The dried cake from the dryers dumps onto the washed coal belt while the effluent discharges to a 10 ft. diameter Roberts & Schaffer flotation machine.

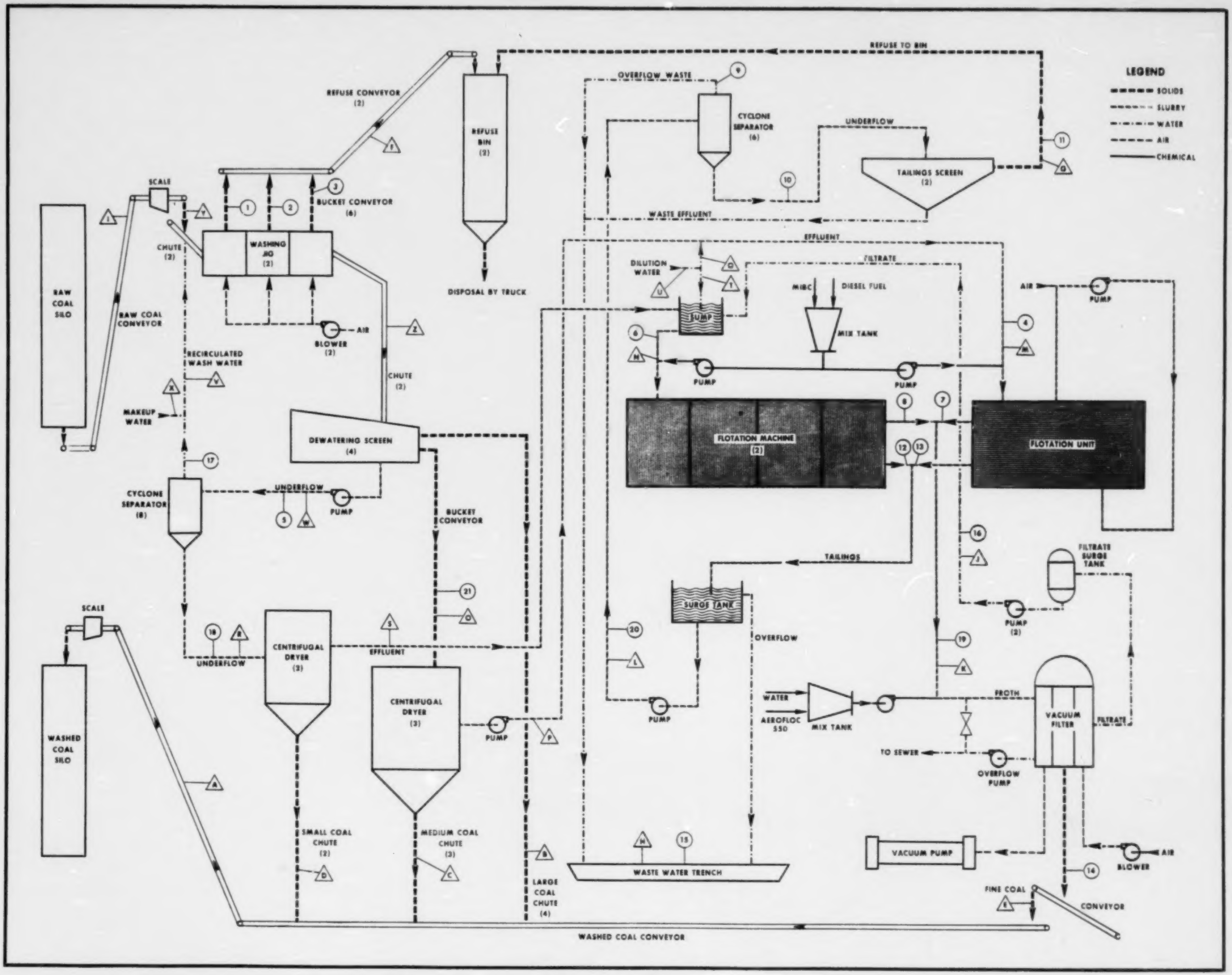
Water and fines from the dewatering screens are pumped to two sets of 16 inch clarifying cyclones. Overflow water is recycled to the jig from which it came. Thickened underflow is fed to two centrifugal dryers from which the cake goes to the washed coal belt and the effluent to two Denver Sub-A cell flotation machines.

In these flotation machines, the fine coal particles contained in the feed from the centrifugal dryers are separated from the slime by air flotation methods and recovered as a low ash froth. The froth from the flotation equipment is then dewatered by vacuum filtration, and discharged as a moist cake to the washed coal conveyor. Associated with the flotation equipment is a tailings disposal system, consisting of six cyclone separators and two tailings screens. The tailings slurry is separated into wet solids and liquid waste. The solid waste passes to the refuse bins. The liquid waste discharges to a waste water trench and then to the coke plant sewer system.

At the current operating rate, the coal washery at the Pueblo Plant processes approximately 4,000 TPD_{10} of raw coal on a seven days a week schedule. The solid waste material amounts to approximately 1,000 TPD_{10}. A high percentage of these solids, approximately 800 TPD_{10}, is separated at the jigs and sent to refuse storage bins. The remaining waste stream passes through a set of cyclones, tailings screens, and on into a settling sump where additional solids are removed before being discharged to the plant sewer. Some 3 to 7 TPD_{10} of the effluent waste solids are coal fines which can be reclaimed and used for coke production.

By Product Coke Plant

After the preparation process the coal is ready to be charged into


LEGEND
SOLIDS
SLURRY
WATER
AIR
CHEMICAL
RAW COAL SILO
RAW COAL CONVEYOR
SCALE
CHUTE (2)
WASHING JIG (2)
BUCKET CONVEYOR (6)
BLOWER (2)
AIR
RECIRCULATED WASH WATER
MAKEUP WATER
REFUSE CONVEYOR (2)
REFUSE BIN (2)
DISPOSAL BY TRUCK
CHUTE (2)
DEWATERING SCREEN (4)
UNDERFLOW
PUMP
CYCLONE SEPARATOR (8)
BUCKET CONVEYOR
CENTRIFUGAL DRYER (2)
CENTRIFUGAL DRYER (3)
UNDERFLOW
EFFLUENT
SMALL COAL CHUTE (2)
MEDIUM COAL CHUTE (3)
LARGE COAL CHUTE (4)
WASHED COAL SILO
WASHED COAL CONVEYOR
REFUSE TO BIN
OVERFLOW WASTE
CYCLONE SEPARATOR (6)
UNDERFLOW
TAILINGS SCREEN (2)
WASTE EFFLUENT
EFFLUENT
FILTRATE
DILUTION WATER
SUMP
MIBC
DIESEL FUEL
MIX TANK
PUMP
FLOTATION MACHINE (2)
FLOTATION UNIT
AIR
TAILINGS
SURGE TANK
OVERFLOW
FILTRATE SURGE TANK
PUMP (2)
WATER
AEROFLOC 550
MIX TANK
FROTH
VACUUM FILTER
FILTRATE
TO SEWER
OVERFLOW PUMP
VACUUM PUMP
BLOWER
WASTE WATER TRENCH
FINE COAL
CONVEYOR
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
20
21
A
B
C
D
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I
J
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Q
R
S
T
U
V
W
X
Y
Z

Figure 1. PROCESS FLOW DIAGRAM COAL WASHERY

coke ovens to be converted into coke. At the CF&I Pueblo Plant (Figure 2) there are four batteries with a total of 206 coke ovens. Total coke production is normally 1800-2400 TPD_{24} with an average of approximately 2000 TPD_{24}. Most of the ovens at the Pueblo Plant are of the Koppers-Becker design (nominal dimensions 40 ft. long, 13 ft. high and 18 inches wide). Production schedules are arranged so that a nominal 20½ hours net coking time is maintained nearly all of the time.

The ovens are the typical slot type where the coal is heated by a regenerative vertical flue system built into the oven walls. Coal is charged through four charging holes in the top of each oven by means of a larry car which travels over the top of the oven batteries. The gases evolved during the coking operation pass along the top of the oven chamber above the coal and out through a pipe at one end into the gas collection system which conveys it to the by-product treatment section. A process flow diagram is presented in Figure 3.

When the coal is completely coked it is ejected from the oven, or "pushed". The doors are removed from each end of the oven, and a coke guide set in place on the discharge side. A car with sloping bottom is spotted at the proper point, and the coke is pushed from the oven by a long ram mounted on the pusher. This ram passes completely through the oven. As the coke is pushed out, the quenching car moves along, thereby spreading the coke out along the bottom of the car. After the ram is withdrawn, the doors are replaced and sealed and another coking cycle is started. The hot coke in the quenching car is taken to the quenching station. There it is sprayed with water and returned to the coke wharf where it is allowed to cool and dissipate moisture.

The composition of the gas evolved during the coking cycle varies greatly. Immediately after charging and heating the coal in the oven, moisture and the small amount of occluded gases present in the coal, mainly oxygen and methane, are driven off. Some carbon dioxide, more methane and the balance of the moisture present will be driven off before the coal "melts" or becomes plastic. The volume of gas evolved during this "heating up" period is small, probably about one percent of the total gas (not including moisture) evolved from the charge.

The major portion of the gas is evolved during fusion of the coal and boiling of the plastic layer. This gas consists mainly of methane and hydrogen. Smaller amounts of carbon monoxide, ethane and higher hydrocarbons are produced. Light oils, consisting of benzene, toluene, xylene and naphthalene, and tar vapors are also produced during this period. The light oils are produced chiefly as a result of "cracking" or breaking down of higher hydrocarbons. This cracking results when the gases and tar vapors evolved from the coal are further heated by passing through the hot coke layer

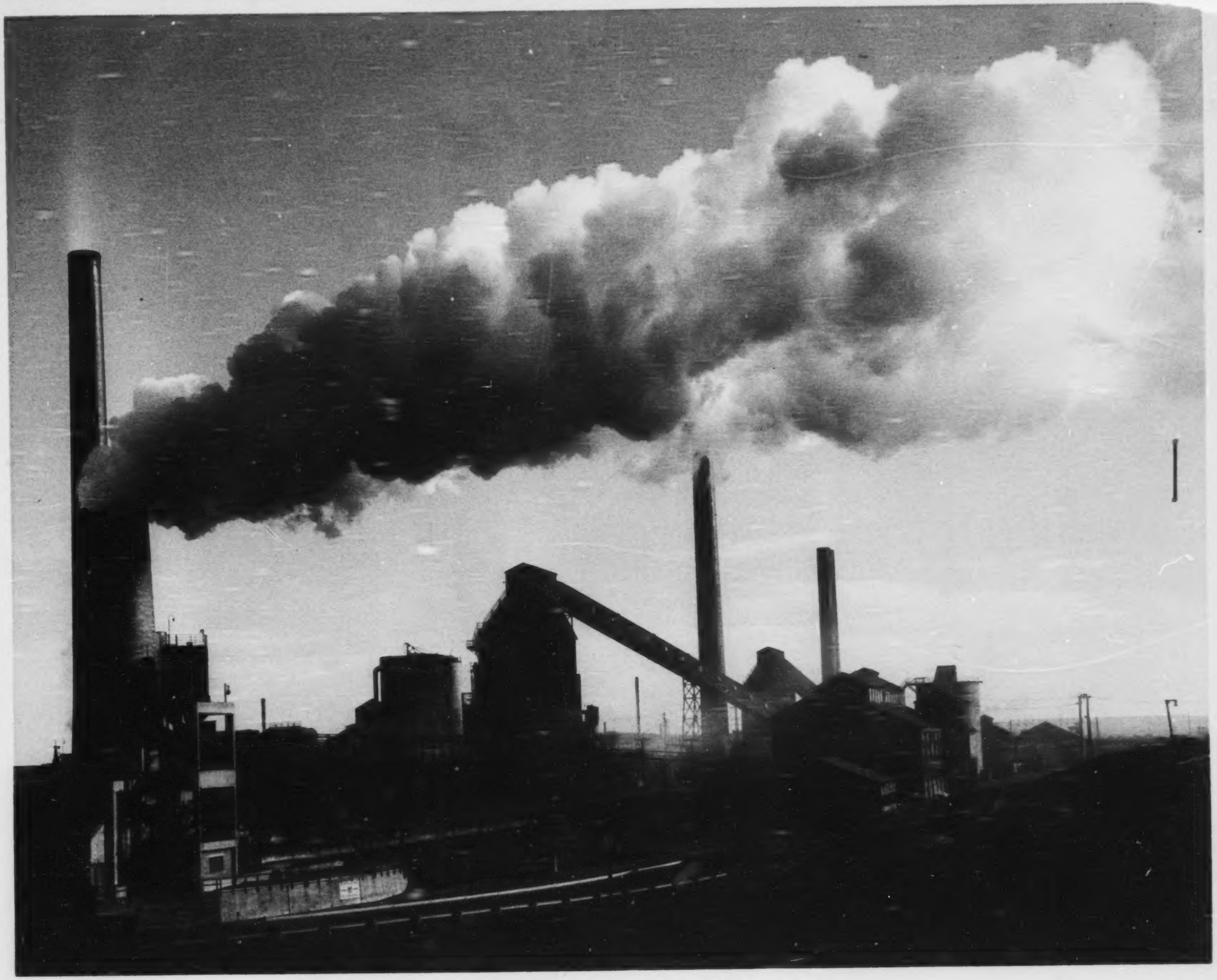


Figure 2. VIEW OF CF&I COKE PLANT

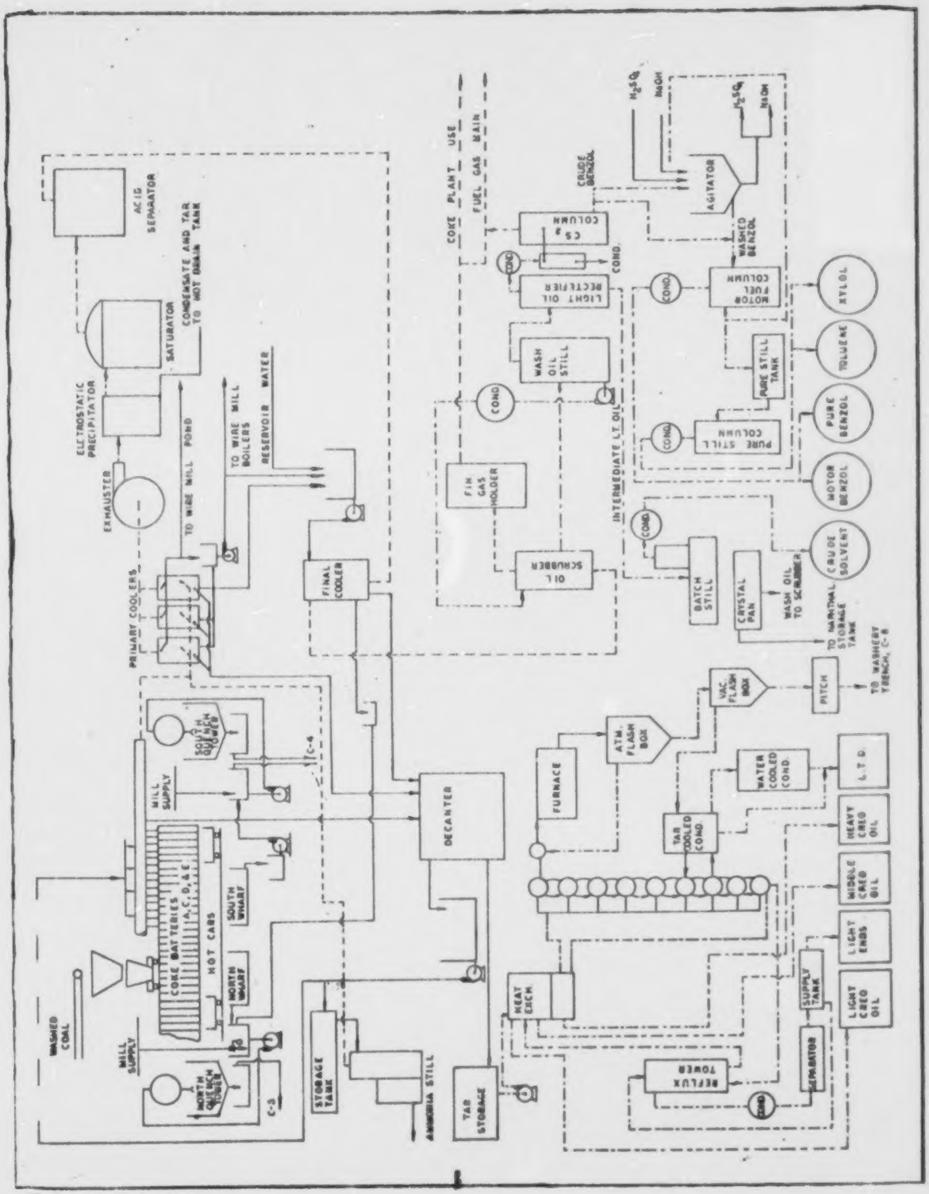

WASHED COAL
MILL SUPPLY
NORTH QUENCH TOWER
C-3
COKE BATTERIES 1 A, C, D, & E
HOT CARS
NORTH WHARF
SOUTH WHARF
MILL SUPPLY
SOUTH QUENCH TOWER
C-4
PRIMARY COOLERS
EXHAUSTER
ELECTROSTATIC PRECIPITATOR
SATURATOR
ACID SEPARATOR
CONDENSATE AND TAR TO HOT DRAIN TANK
TO WIRE MILL POND
TO WIRE MILL BOILERS
RESERVOIR WATER
FINAL COOLER
DECANTER
STORAGE TANK
AMMONIA STILL
TAR STORAGE
HEAT EXCH.
FURNACE
ATM. FLASH BOX
VAC. FLASH BOX
PITCH
TO WASHERY TRENCH, C-8
TAR COOLED COND
WATER COOLED COND.
L.T.D.
HEAVY CREO OIL
MIDDLE CREO OIL
LIGHT ENDS
LIGHT CREO OIL
SUPPLY TANK
SEPARATOR
REFLUX TOWER
COND
FIN GAS HOLDER
OIL SCRUBBER
WASH OIL STILL
INTERMEDIATE LT. OIL
LIGHT OIL RECTIFIER
CS2 COLUMN
COND.
COKE PLANT USE
FUEL GAS MAIN
CRUDE BENZOL
BATCH STILL
CRYSTAL PAN
WASH OIL TO SCRUBBER
TO NAPHTHALENE STORAGE TANK
CRUDE SOLVENT
MOTOR BENZOL
PURE BENZOL
TOLUENE
XYLOL
PURE STILL COLUMN
PURE STILL TANK
MOTOR FUEL COLUMN
WASHED BENZOL
AGITATOR
H2SO4
NaOH

Figure 3. PROCESS FLOW DIAGRAM BY-PRODUCT COKE PLANT

and along the hot oven walls. Upon leaving the oven the gas passes into a collection system and through the by-product recovery sections of the coke plant.

The gas passes from the oven through the standpipe into the collecting main at a temperature of about 750°F. In the collecting main the gas is sprayed with flushing liquor and cooled to about 175°F. This cooling condenses about 75% of the tar vapors and most of the moisture in the gas. The condensed water is saturated with ammonia and is called ammonia liquor or flushing liquor. The condensed liquor and tar flows to a decanter tank where the tar and liquor are separated. The tar is pumped to a tar storage tank; the liquor needed for flushing is pumped back to the collecting main sprays and the excess is pumped to storage. The excess can be put through ammonia stripping stills where low pressure steam "strips" the ammonia from the liquor. If used, the stripped ammonia is returned upstream of the primary coolers and the stripped liquor sent to sewer.

The gas from the ovens flows through the suction mains to the indirect type of primary coolers where it is further cooled to about 85°F. This cooling results in the condensation of most of the remaining tar and some of the moisture so that the gas going to the exhausters is saturated with moisture and carries some tar vapors in the form of a fog or mist. From the time the gas leaves the collecting main until it gets to the exhausters it is under negative pressure ranging up to a maximum of about 400mm of water. After the exhausters the gas is under positive pressure ranging up to a maximum of 70 inches of water. The hot uncontaminated cooling water is mixed with fresh mill water in a concrete basin and re-used for roll cooling at the Rod Mill. Approximately 5% of the total tar in the gas remains after the primary coolers. Some of this tar is removed by impingement on the vanes of the centrifugal blowers or exhausters as the gas is forced through the system. After the exhausters, the gas passes through electrostatic precipitators where the final traces of tar are removed by electrostatic action and returned to the decanter. The production of tar amounts to about 11 gallons per ton of coal carbonized or about 11,000,000 gallons per year.

CF&I coal, when coked, produces about 5 pounds of ammonia per ton of coal carbonized. Approximately 85% of this ammonia remains in the gas through the cooling operation; the remainder is removed from the gas in the form of ammonia liquor.

The gas stream, after tar removal, is passed through the saturator or ammonia absorber. Known as the semidirect process for ammonia recovery, the ammonia is removed from the gas by reaction with a weak sulfuric acid bath. The resultant ammonium sulfate slurry is used in the manufacture of fertilizer.

After the removal of ammonia from the gas in the saturators, the gas still contains light oils and naphthalene. These are removed before distributing the gas as fuel gas. When the gas comes from the saturators it is under a positive pressure of 30 to 40 inches of water and is about 125°F. In preparation for light oil recovery, the gas is cooled to about 80°F in a direct water spray cooler referred to as a final cooler. Cooling the gas by direct water spray condenses some of the naphthalene from the gas. This naphthalene is entrained in the water spray. The water, before leaving the final cooler, flows through a tar bath which washes the naphthalene from the water. The tar in the final cooler is constantly being charged with fresh tar. The naphthalene enriched tar is returned to the decanter. A portion of the water from the final cooler is used for cooling the hot coke pushed from the ovens.

After the gas has been cooled and some of the naphthalene removed, it is ready for light oil removal in the light oil scrubbers or absorbers. These scrubbers are packed columns with two or more operated in series. The gas flows upward through the packing. Absorber oil (a petroleum oil of about the consistency of light machine oil) flows downward through the packing. The absorber oil collects or absorbs the light oil and the remaining naphthalene from the gas stream. After removal of the light oil, the gas is distributed for fuel gas. About 45% is returned for under firing the coke ovens and about 55% is used as fuel for the rolling mills.

The enriched absorber oil coming from the scrubbers is pumped to the benzol plant where it is heated. The light oil is removed from the absorber oil by steam stripping. The stripped or "debenzolized" oil is cooled and recycled to the scrubbers for absorption of more light oil. The uncontaminated cooling water is mixed with the primary cooling water discharge.

After being stripped from the absorber oil, further separation into secondary light oil and intermediate light oil takes place in the rectifier. The secondary light oil contains crude benzol, toluol and xylol. The intermediate fraction contains crude solvent, naphthalene and absorber oil.

The secondary light oil flows to a continuous fractionating column. Here the light or fore-runnings compounds (consisting mainly of carbon di-sulphide and cyclopentadiene) are vaporized and fed back into the fuel gas main. The bottoms from the CS_2 column contain crude benzol, toluol and xylol less the fore-runnings compounds. These flow to storage. The next step in purification of the secondary light oil is a wash with sulphuric acid in the agitator. This operation removes the last of the sulphur compounds and the so-called unsaturated compounds. Spent acid is removed from the agitator by decanting and is reused at the saturator. Following acid treatment, light oil is neutralized with sodium hydroxide. The washed BTX (benzol, toluol, xylol) is distilled in the motor fuel or benzol fractionating column. Pure benzol is produced. The remaining TX (toluol and xylol) fraction is distilled in a batch still which separates toluol and xylol. The residue from this distillation is shipped to the Open Hearth for fuel.

The intermediate light oil is fractionated in a batch still. The crude solvent is distilled off and stored for sale. The bottom product, containing naphthalene and absorber oil, is cooled in large pans. This cooling permits the naphthalene to crystallize. The oil is drained from the pans and returned to the absorber oil system. The naphthalene is melted and transferred to crude naphthalene storage tanks and either sold as crude naphthalene or used at the Open Hearth as fuel, depending on market conditions.

Tar recovered from the decanter is processed at the Creosote Plant. A portion of the tar produced at the plant is processed at the Creosote Plant to meet market requirements. Excess tar is shipped to the Open Hearth for fuel. In the Creosote Plant the tar is heated. The creosote is vaporized and is partially condensed into various creosote fractions. The residue from the distillation is a hard pitch which is used in the manufacture of roofing pitch. The excess is charged back into the ovens with the washed coal.

The normal ranges for products per ton of dry coal carbonized are as follows:

Principal Products	Quantity (Lb/Ton) of Dry Coal
Coke (Lb)	1,300-1,550
Gas (Ft^3)	10,000-12,500
Crude Tar (Gal.)	5-12
Ammonia as NH_3 (Lb)	4-7
Crude Light Oil (Gal.)	2-4
Naphthalene (Lb)	1.6

Theoretical Approach and Project Design

The description of the coal washery and by-product coke plant facilities discloses many potential and actual sources of water pollution. As a first step in eliminating water pollution from the coke plant area, CF&I designed and conducted a research program as follows:

Waste and Sources Survey

The initial phase of the project provided for a detailed study of the coal cleaning, coking, and coal chemical recovery operations to determine all sources of water pollution. All waste water flows originating in the coke plant area, continuous or intermittent, were to be sampled and analyzed to determine physical and chemical characteristics.

The physical parameters measured included flow, temperature, total solids, total suspended solids, fixed solids and total fixed solids. The chemical parameters measured included:

pH
Ether Extractables
Total Hardness
Alkalinity
Chloride
Nitrate
Sulfate
Ammonia-Nitrogen
Phenol
Cyanide
COD
BOD_5
Heavy Metals (Cu, Zn, Pb, Fe, Mn, Ni on total Coke Plant effluent only)

The information gathered in this phase of the study described the contaminated waste flows and provided data for selecting the type of bench scale treatment processes to be tested under the second phase of the study.

Treatability Studies

This phase of the study was designed to determine applicable methods and costs of isolating, eliminating and/or disposing of contaminants for a single or multi-stage operation. A number of mechanical, chemical and biological treatment processes were selected for the tests. A literature survey was made during the project period to provide new process developments applicable to the treatment of the coke plant waste.

Those treatment processes selected for actual laboratory and bench scale study were:

Biological Oxidation - Two biological systems, activated sludge and the trickling filter, were selected for study. The acclimation of a bacterial cell mass to phenol and cyanide was a prerequisite. Plans for the activated sludge study included primary settling, aeration basins and secondary settling. A literature review was planned to determine the feasibility of using oxygen or ozone as an oxygen source for the activated sludge system. The trickling filter system design included primary settling, automatically-controlled flow and a re-cycle system.

Adsorption - Adsorption by activated carbon for removal of color, phenols and other organic compounds was another waste treatment process selected for study. In conjunction with this study, samples of CF&I fine coal and coke breeze were tested for adsorption capabilities for comparison against the activated carbon. A literature review was planned for determining the methods and costs of regenerating exhausted activated carbon.

Chemical Treatment and Clarification - A chemical treatment investigation was designed to study the conversion of fixed ammonia to free ammonia and the removal of color by addition of lime or caustic soda. These studies were to coincide with the steam and air stripping of ammonia. Cost comparisons were made between hydrated lime and the calcined lime produced at CF&I.

Clarification studies included solids settling rates and methods of suspended solids removal such as high rate filtration.

Steam and Air Stripping of Ammonia - The operational parameters of stripping ammonia by steam were studied on a bench scale in preparation for performance testing of one of the two existing ammonia stills.

Stripping of ammonia by air was included in order to obtain a comparison in removal efficiencies and operating costs between steam and air stripping.

Ion Exchage - Various ion exchange resins were tested and compared for reductions in ammonia, phenol, and cyanide concentrations, volumes of waste treated before breakthrough and methods of regenerating the resin following breakthrough.

Filtration and Sedimentation of Coal Washery Wastes - This study consisted of three phases. Jar tests determined the washery waste settling rates, the effect of organic and inorganic coagulents on settling rates and the effect of pH adjustment, especially with waste pickle liquor, on settling rates. High rate filter performance was examined coincidental with experiments to determine the effect of additives on filter efficiency and rate. A third phase tested waste water re-use possibilities. This included vacuum filtration for sludge dewatering, large scale settling tests to obtain sizing information for concentrating waste water sludge, and determination of the re-cycle effect on water quality.

Process Modification and Improvement - Present operating and maintenance practices were reviewed to determine improvements and changes that will reduce pollutant concentrations that are being discharged to the sewer from normal and abnormal operations.

Study of Other Treatment Methods - A review of the literature was conducted during the course of the project with emphasis placed on other possible treatment methods such as reverse osmosis, electrodialysis and the use of oxygen or ozone as an oxygen source in biological treatment processes. Any treatment process which appeared to have application was bench tested.

Solid Waste Investigations - Investigation into the handling of solid wastes from the coal washing operation was started by determining the variations of gangue material in the raw coal supply from the mines. Following this, previously planned improvements were evaluated and compared with alternative methods of fine coal recovery to prevent loss of fine coal to the sewer.

Selected methods for treating the waste material remaining after fine coal recovery were correlated with the results of the Filtration and Sedimentation phase of the study and the results of additional studies of alternative methods.

Disposal methods of solid wastes which were evaluated included trucking the wastes to a dump area as is presently done, pumping the wastes to a dump area, and marketability for part or all of the gangue material.

Design and Engineering Coal Washery

Design drawings were made of the improvements necessary to prevent the loss of fine coal to the sewer and the system required to handle waste material from the coal washing operation. Data collected during the "Filtration and Sedimentation" and the "Solid Waste Investigation" studies conducted under this project provided the necessary information for preparing equipment specifications. System cost estimates were prepared from the engineering specifications of Stearns-Roger Corporation and from proposals made by various equipment manufactures.

WASTE AND SOURCES SURVEY

The research study of the Coal Preparation Plant and the By-Product Coke Plant actually started with the development of a sampling program that would provide chemical and physical characteristics of all liquid flows originating in the coke plant area. Drawings of the sewers in the area were obtained and individual manholes systematically numbered to provide a method for recording the data collected. In addition, interviews were held with coke plant operating and maintenance personnel to make sure that all existing drain line flows, not shown on the sewer drawings, would be included in the sampling program.

Methods of Sampling

Grab and composite samples were taken from all flows enumerated in the sampling program. Temperature and flow quantities were determined and recorded at the time of sampling. Sample bottles containing preservative agents for ammonia, cyanide, phenol, sulfide, and COD were prepared prior to sampling.

Chemical Analysis, Procedures

Analyses for ammonia, total suspended solids, phenolphthalein alkalinity, total alkalinity, chloride, total hardness, pH, and conductance were made both on the grab samples and on composite samples in order to establish loads. Temperature and conductivity recorders were placed in some of the more critical discharges such as the effluent from the quenches, the total plant effluent, and the two principal branches of the sewer system at points C-18A and C-7B (See Figure 4).

With the exception of the coal washery, the plant operates at a reasonably uniform rate; therefore, most of the composite samples were collected on the first and second shift. A minimum of 4 grab samples at 1½ hour intervals were used in the preparation of composite samples.

Additional analyses of ortho phosphate, nitrate, total dissolved solids (gravimetric), percent of volatiles on total dissolved solids, total organic carbon, and carbon tetra chloride extractables were made on some samples before and after start-up of the ammonia still.

Analytical procedures and sample presentation followed Standard Methods for the Examination of Water and Wastewater, 12th Edition, 1965. The only exceptions were procedures used in the analyses for

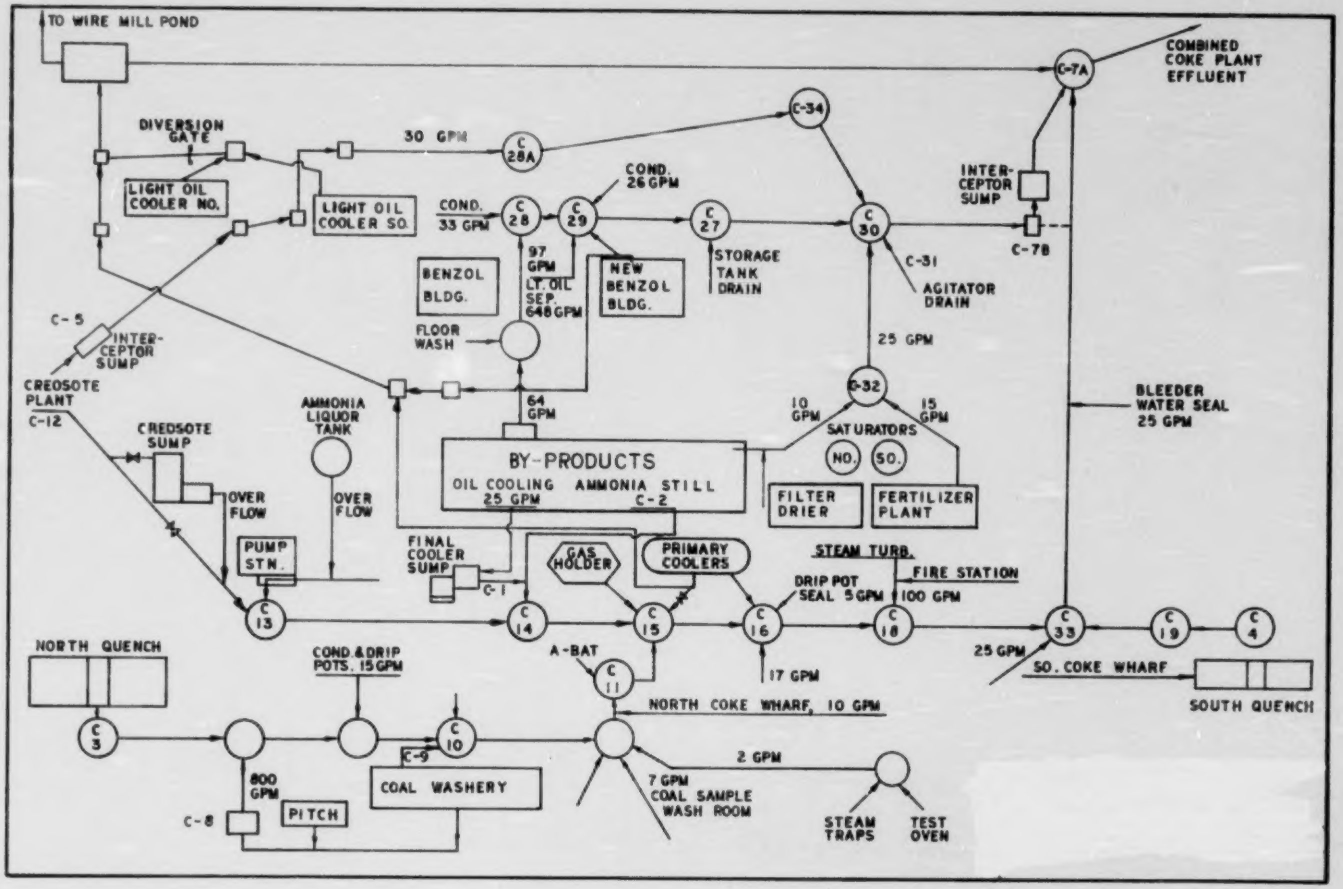

TO WIRE MILL POND
COMBINED COKE PLANT EFFLUENT
C-7A
C-34
DIVERSION GATE
30 GPM
C 28A
LIGHT OIL COOLER NO.
LIGHT OIL COOLER SO.
COND. 33 GPM
C 28
C 29
COND. 26 GPM
C 27
C 30
INTER-CEPTOR SUMP
C-7B
97 GPM
LT. OIL SEP. 648 GPM
BENZOL BLDG.
NEW BENZOL BLDG.
STORAGE TANK DRAIN
C-31
AGITATOR DRAIN
C-5
INTER-CEPTOR SUMP
FLOOR WASH
25 GPM
CREOSOTE PLANT C-12
64 GPM
C-32
10 GPM
15 GPM
BLEEDER WATER SEAL 25 GPM
CREOSOTE SUMP
AMMONIA LIQUOR TANK
SATURATORS
NO.
SO.
BY-PRODUCTS
OIL COOLING 25 GPM
AMMONIA STILL C-2
OVER FLOW
OVER FLOW
FILTER DRIER
FERTILIZER PLANT
PUMP STN.
FINAL COOLER SUMP
C-1
GAS HOLDER
PRIMARY COOLERS
STEAM TURB.
FIRE STATION
DRIP POT SEAL 5 GPM
100 GPM
C 13
C 14
C 15
C 16
C 18
C 33
C 19
C 4
25 GPM
NORTH QUENCH
A-BAT
C 11
17 GPM
SO. COKE WHARF
SOUTH QUENCH
COND. & DRIP POTS. 15 GPM
NORTH COKE WHARF, 10 GPM
C 3
C 10
C-9
2 GPM
800 GPM
COAL WASHERY
C-8
PITCH
7 GPM COAL SAMPLE WASH ROOM
STEAM TRAPS
TEST OVEN

Figure 4. SEWER FLOW SHEET

heavy metals and total organic carbon. A Beckman Model 915 Total Organic Carbon Analyzer was used for total organic carbon. Spectrographic and atomic absorption methods were used in determining heavy metals. All procedures are referenced in Appendix H, p. 268.

In conjunction with the chemical analyses, a program was initiated to determine the precision and accuracy of some of the analyses, namely phenol, cyanide and ammonia nitrogen. This program was adopted from the Laboratory Quality Control Manual published in 1969 by the Federal Water Pollution Control Administration. The program was especially helpful in training several of the CF&I chemists assigned to the study. The results are presented in Appendix I, p. 270. Additional cross-checks of analytical results through the use of duplicate samples were obtained during the study of adsorption by activated carbon. Duplicate samples were analyzed by CF&I chemists and the Denver Laboratory of Calgon Corporation.

Methods of Flow Measurements

Various flow measurement techniques were employed dependent upon conditions at the sampling point. The measuring techniques were stream velocity, liquid height at an overflow weir, salt injection, and fill time in a bucket.

Velocity readings are meaningful flow measurements only when the stream runs with minimum turbulence. The bottom of the sewer or outfall must either be completely clean or be covered with a blanket of sludge which is deposited at an even thickness and is free of other debris. If the sewer bottom is broken or contaminated with bricks, rocks, or similar material, the velocity reading is not representative of the actual flow because of eddy currents and turbulence.

A velocity meter, manufactured by Gurley Hydrological Instruments, Troy, New York, was used for velocity measurements. The meter was a Gurley Model No. 622-F which had previously been calibrated by Rensselaer Polytechnic Institute. Actual flow rates were calculated using velocity readings and the area of wetted perimeter in the sewer.

Flow measurement by salt injection was used in sewers where current velocity readings were meaningless. The salt injection method is based on a material balance of salt injected into the sewer at a constant rate. Complete mixing of the salt solution and the sewer flow is necessary between the point of injection and the point of sampling.

Flow rate was calculated as follows: $Q_1 = \dfrac{Q_2 C_2}{C_3 - C_1}$

Where: Q_1 = Flow rate in sewer (gal/min)
Q_2 = Rate of addition of NaCl solution (gal/min)
C_1 = NaCl concentration in waste stream above point of addition (mg/l)
C_2 = NaCl concentration in solution added (mg/l)
C_3 = NaCl concentration in waste stream below point of addition (mg/l)

More direct methods of flow measurements were made in those cases where the stream could be made to overflow a weir or dam. Standard handbook tables were used for calculation of weir overflow rates.

Small streams were sized by timing the fill rate in small containers such as a five gallon bucket. Such techniques are possible only when the stream discharge is at an elevated position. This method was used primarily for condensate discharges, drip pots, and seal water.

A summary of observed flow rates for all sampling points is shown in Table 1 along with maximum and minimum values for most of the larger flows. Temperatures are also included in this table to make temperature comparisons of the plant effluent before and after diversion of uncontaminated streams.

The flow rates also appear on the general sewer flow diagram, Figure 4.

It is evident from Table 1 that the largest flows are found at the quench stations, the coal washery, and the light oil separator in the benzol plant. The discharge from the benzol building (744 gpm) is a mixture of condenser cooling waters which is not contaminated.

Waste Stream Characterization

Sample Point C-1 This is the overflow from the final cooler. It also receives cooling water from gas booster pumps (flow rates variation is 25-200 GPM). The final cooler effluent is used for coke quenching and normally there is no overflow from the sump to the sewer.

Table 2 shows the analytical results of three grab samples collected in January and in June of 1969. The data indicate that final cooler overflow (used for coke quenching) has high levels of alkalinity, TDS, and COD.

In addition to these data, CF&I records which date to April of 1961 indicate that the final cooler effluent also contains several hundred mg/l of ammonia (maximum recorded 1216 mg/l, minimum 75 mg/l). Cyanide values have ranged from 28-200 mg/l, although the most common concentrations are between 100 and 200 mg/l. Generally phenol concentration is relatively constant at about 50 mg/l; however, concentrations have exceeded 200 mg/l.

TABLE 1

Summary of Flow and Temperature Data from Coke Plant Study

Stream to Sampling Point	Location Operating Area	Flow Norm.	gpm Max.	Min.	Temp °C
C-1	Oil Cooling	25	-	-	21
C-3	No. Quench	500	700	200	71
C-4	So. Quench	300	500	200	71
C-8	Coal Washery	600	700	500	21
C-11A	Cond. & Drip Pots	15	-	-	54
C-11A	Steam Traps	11	-	-	83
C-11	No. Coke Wharf	10	-	-	43
C-13	Creosote Plant	0	30	0	-
C-2	Ammonia Still	75	150	50	93
C-16	Primary Cooler	75	150	50	18
C-16	Drip Pot	5	-	-	54
C-16	Drip Pot	17	-	-	54
C-18	Steam Turbine	100	-	-	54
C-33	Pot Drain	25	-	-	54
C-7A	Bleeder Water	25	-	-	54
C-5	Interceptor Sump	30	-	-	-
C-28	Condensate	33	-	-	54
-	By-Products Plant	64	-	-	-
C-29A	Light Oil Sep.	114	150	100	38
C-29A	Benzol Bldg.	744	800	500	46
C-29A	Condensate	26	-	-	83
C-30	Saturator Area	25	-	-	-
	Est. Misc.	150	200	100	-
	TOTAL	2969	3380	1700	

REMARKS:

Calculated average temp point C-7A = 45 °C
From Fig. 5, Max. Temp C-7A 46.6 °C, min 24 °C
Approximately average, 36 °C (estimate and convert to °C From Fig. 5)

TABLE 2

Sampling Points
C-1, C-2, C-5, C-12

	Final Cooler	Ammonia Stills			Creosote Sump		Creosote Recycle	
Sample Source	C-1*	C-2*	C-2*	C-2*	C-5	C-5	C-12	C-12
Date	7-7-69	3-3-69	6-15-69	7-7-69	7-7-69	7-16-69	7-3-69	7-16-69
Time of Day	8:00 AM	8:00 AM	1:30 PM	--	--	--	--	--
Type of Sample	Grab	Grab	Grab	Grab	Grab	Grab	Grab	Grab
Temp., ^{o}C	30	23	--	29	43	43	43	--
pH	8.7	8.3	9.1	8.4	7.4	7.4	7.4	--
P-Alk., $CaCO_3$ mg/l	266	6	108	44	0	--	--	--
M-Alk., $CaCO_3$ mg/l	950	142	--	192	94	--	--	--
Chloride, NaCl mg/l		30	12	6	2	--	--	--
Tot. Hard., $CaCO_3$ mg/l	84	198	130	130	120	--	--	--
Sulfate, SO_4 mg/l	42	--	--	63	21	--	--	--
Nitrite, NO_2 mg/l	1.2	--	--	$<$0.1	2.2	--	--	--
TDS (grav) mg/l	1170	--	222	203	164	--	--	--
Ammonia, NH_3 mg/l	--	8.5	--	--	0	0	0	4.3
Cyanide, CN^- mg/l	--	1.2	--	--	--	.3	.8	.4
Phenol, C_6H_5OH mg/l	--	2	--	--	--	110	33	136
COD, Oxygen Demand mg/l	1735	77	--	113	--	--	--	--
Ether Extr. mg/l	--	10	--	--	--	--	--	--
TSS mg/l	174	54	161	143	108	--	--	--

Remarks: *Final cooler sump overflowing

Ammonia still not operating

Conductance and temperature recorder charts revealed no significant variation in these parameters. Conductance ranged from 1300-2600 μ mhos/cm and temperatures varied between 49-57 °C.

Sample Point C-2 represents the effluent from the ammonia still (when the ammonia still is operating) and accumulated leakage from valve packings and condensate water in the primary cooler and ammonia still area. A flow rate of 50 gpm was observed. The ammonia still which is not operating at this time would increase this flow to 125 gpm. Contaminates would be ammonia, phenol, and cyanide. Analyses of sample point C-2 are found in Table 2.

Following start-up of the ammonia still about November 5, 1969, continuous discharge of weak ammonia liquor has been sampled at this point. Analyses from the still operation are found in Tables 65-66, page 207-208.

Sample Points C-3 and C-4 These represent quench water discharges. The quench water is a mixture of water from the final cooler sump and fresh mill makeup water.

The final cooler water spills over an 18" weir in the cooler sump before going to the quench stations. Flow rate through the final cooler is regulated manually to control the temperature of the exit gas. Variations in temperature of the gas result in large water flow fluctuations of 250-800 gpm.

Normally the water from the final cooler is split to the two quench stations in equal flow; however, the north quench is used more frequently than the south. The overflow from the north quench, therefore, is normally intermittent while the south quench overflows practically all the time.

The north quench system uses approximately 6000 gallons/quench, of which 2200 gallons evaporate. Water use at the south quench varies with the load in the quench car. Coke loads from the "E" battery require approximately 5500 gallons/quench. Loads from other batteries may need as little as 3500 gallons/quench.

Tables 3-5 contain analytical data.

The pH values obtained from the two quenching stations do not reflect any difference in water use. Alkalinity and chloride results obtained from the two quenching stations indicate that there is a buildup of these solids during the quench since the data shows generally higher concentration in the north quench overflow. Values for ammonia, cyanide, phenol, COD, sulfide, ether extractables and total suspended solids show some variation with ammonia, cyanide and phenols generally higher in the north. The cyanide concentration in the north quench appears to be somewhat more constant than in the effluent from the south quench. South quench range in concentration is <1 to 68 mg/l. Normal cyanide concentrations are less than 10 mg/l in the south quench effluent water

TABLE 3

SAMPLING POINTS C-3 & C-4

NORTH AND SOUTH QUENCH

DATE	TIME TAKEN	TYPE OF SAMPLE	TEMP. °C	pH	CONDUCTIVITY µ mhos/cm	AMMONIA NH_3 Mg/1	TSS Mg/1	TOC Mg/1	COD Mg/1	BOD Mg/1
North Quench										
2-24-69	8:15am	Grab	60	8.5		21.6	120		813	-
6-16-69	8:00am	"	-	-		-	-		212	38
7-7-69	9:05am	"	57	8.9		-	198		357	-
11-26-69	8:00am	"	58	8.5		1.7*	130			
	9:00am	"	60	8.5		2.1*	65			
	10:00am	"	55	8.5		1.8*	79			
	11:00am	"	54	8.6		2.0	24			
	12:00pm	"	59	8.6		0.8	170			
11-27-69	8:30am	"	36	8.3		3.0*	168			
	9:30am	"	36	8.3		.2*	145			
	10:30am	"	57	8.5		0.8	127			
	11:30am	"	60	8.6		1.7*	190			
	12:30pm	"	58	8.6	400	3.0*	50	11.6		
12-1-69	8:00am	"	57	8.5		2.0*				
	9:00am	"	58	8.5		4.1*				
	10:00am	"	56	8.6		4.1*				
	11:00am	"	58	8.5		2.4*				
	12:00pm	"	55	8.5	440	5.7*				
South Quench										
1-30-69	8:15am	Grab	28	8.2		-	45		-	-
2-24-69	8:25am	"	19	7.6		34.4	19		864	-
5-16-69	8:00am	"	-	-		-	-		310	48

Remarks: *By Distillation at pH 9.4

TABLE 4

Sampling Point C-3

North Quench

Date	11-26-69	11-27-69	12-1-69
Time Taken	8A - 12P	8:30A-12P	8A - 12P
Type of Sample	Comp	Comp	Comp
P-Alk, $CaCO_3$ mg/1	6	0	4
M-Alk, $CaCO_3$ mg/1	76	64	70
Chloride NaCl mg/1	3	16	2
Tot. Hard. $CaCO_3$ mg/1	142	142	138
Sulfate, SO_4 mg/1	163	163	163
O-Phosphate, PO_4 mg/1	0	0	0
Nitrate, NO_3 mg/1	0	0	0
TDS (Grav) mg/1	260	260	250
Cyanide, CN mg/1	<.01	<.01	<.01
Phenol, C_6H_5OH mg/1	.21	.14	.27
Ether Extr. mg/1	2	4	1

Cyanide and Phenol Analyses

12/2/69 - 12/11/69

Time	2:00P	3:00P	4:00P	5:00P
CN	24	20	21	19
C_6H_5OH	16	16	20	14
Time	10:00A	2:00P	4:00P	
CN	22	14	5	
C_6H_5OH	15	15	.11	
Time	2:15P	5:00P		
CN	15	14		
C_6H_5OH	23	20		
Time	8:30A	11:00A	1:00P	5:30P
CN	20	16	7	22
C_6H_5OH	19	24	25	32
Time	8:30A	12:30P	5:00P	
CN	8	13	6	
C_6H_5OH	30	27	29	
Time	8:30A	1:00P		
CN	3	15		
C_6H_5OH	22	38		

TABLE 5

Sampling Points C-3 - C-4

Comparison of North and South Quench

Date	2-24-69	3-4-69	3-5-69	3-6-69	3-13-69	3-19-69	7-7-69
Time of Day	8:15 AM	1st sh	1st sh	2nd sh	2nd sh	3rd sh	9:05 AM
Type of Sample	Grab	Comp	Comp	Comp	Comp	Comp	Grab
Sampling Freq.	Once	2 Hrs	2 Hrs	2 Hrs	2 Hrs	2 Hrs	Once
Temp Range °C	60	60	-	63-65	69-73	62-65	57
pH	8.5	8.7	8.5	8.8	8.9	8.6	8.9
Cond. μ mhos/cm	-	800	550	-	-	-	-
Min-Max	-	-	420-1120	-	-	-	-
P-Alk, $CaCO_3$ mg/1	36	40	8	80	110	62	250
M-Alk, $CaCO_3$ mg/1	142	138	90	230	330	216	460
Chloride, NaCl mg/1	24	60	70	40	90	100	16
Tot Hard. $CaCO_3$ mg/1	196	158	166	120	164	130	94
Sulfate mg/1	-	-	-	-	-	-	67
TDS (grav) mg/1	-	-	-	-	-	-	488
Ammonia, NH_3 mg/1	21.6	32.7	12	89	85	61.6	-
Cyanide, CN mg/1	20.6	11.3	14.3	16.8	18.5	20.4	-
Phenol, C_6H_5OH mg/1	24	6	17	25.8	20.2	15.5	-
COD mg/1	813	196	316	382	184	776	357
Sulfide mg/1	-	2	1	5.8	1.5	10.2	-
Ether Extr. mg/1	10	18	59	5	-	2.5	-
TSS mg/1	120	409	317	371	504	3060	198

NORTH C-3

Date	1-30-69	2-24-69	3-4-69	3-5-69	3-6-69	3-13-69	3-19-69
Time of Day	-	8:25 AM	1st sh	1st sh	2nd sh	2nd sh	3rd sh
Type of Sample	Grab	Grab	Comp	Comp	Comp	Comp	Comp
Sampling Freq.	Once	Once	2 Hrs	2 Hrs	2 Hrs	2 Hrs	2 Hrs
Temp Range °C	28	19	49-60	27-60	49-63	54-68	59-62
pH	8.2	7.6	8.3	8.6	8.5	8.9	8.4
Cond. μ mhos/cm	-	-	-	-	940	940	-
Min-Max	-	-	-	-	350-1600	330-1300	-
P-Alk, $CaCO_3$ mg/1	46	0	10	42	20	90	10
M-Alk, $CaCO_3$ mg/1	264	186	80	140	100	240	82
Chlorides, NaCl mg/1	17	24	60	70	40	60	50
Tot Hard. $CaCO_3$ mg/1	-	200	158	166	168	190	160
TDS (grav) mg/1	465	-	-	-	-	-	-
Ammonia, NH_3 mg/1	-	34.4	5.5	6	12	51.4	7.7
Cyanide, CN mg/1	-	68.8	3.8	3.4	1	10.6	.1
Phenol, C_6H_5OH mg/1	-	25.3	1.5	1.9	1	8	.2
COD mg/1	-	864	278	482	159	211	111
Sulfides mg/1	-	-	0	.5	1.5	12.5	1.5
Ether Extr. mg/1	-	16.6	13	8	7.3	-	14.5
TSS mg/1	45	19	428	307	434	860	210

SOUTH C-4

and 15 to 20 mg/l in the north quench water. The phenol concentrations in the south quench discharge also show a tendency to be lower than in the north quench. Normal levels are less than 5 mg/l in the south quench and less than 30 mg/l in the north quench. No specific pattern can be seen from the COD, ether extractables and total suspended solids data.

The heavier use of the north quench is indicated by the relatively steady temperature of its waste discharge. The periodic drop in temperature at the south quench is due to the use of fresh water. Conductance also drops when there is no quenching.

Sample Point C-5 This is the discharge from the interceptor sump in the Creosote Plant area which catches tar spilled from operations at that plant. The tar settles to the bottom of the sump and is reclaimed by pumping the collected tar to storage. Steam traps and wash hoses continuously flow into the spilled tar collection trenches at the rate of 30 gpm. Contaminants are phenolic compounds and cyanides. Any tar escaping the interceptor sump is recovered in the Benzol yard interceptor sump located further downstream. The data obtained from this sampling point are shown on Table 2.

Sample Point C-7A This is the combined flow from the Coke Plant, including the coal washery, by-product and creosote plants. In addition, cooling water from the primary coolers is diverted to this point when water is not being repumped to CF&I's Rod Mill. Flows, not including the cooling water from the primary coolers, vary between 2000 gpm and 3500 gpm. Primary cooler flow varies from 1700 gpm to 2500 gpm.

Based on the data shown in Tables 6-8 and on earlier work done by CF&I, the temperature of the combined coke plant effluent has apparently been reduced considerably between 1967, when the average temperature was 44-46°C., and 1969. The 1969 data indicates that some seasonal variation may be expected. Figure 5 reflects typical daily variation. The pH of the combined effluent is fairly steady, generally between 8.5 and 9.0. Considerable variation in ammonia is experienced at this point. However, concentrations above 500 mg/l are seldom encountered.

Cyanide concentrations vary also. A total of 33 grab samples tested earlier include 5 having concentrations above 50 mg/l. The 12 composite samples collected in March, 1969, have an average concentration of 26.7 mg/l cyanide. Higher cyanide levels occur occasionally as seen in the grab samples collected from 1964 to 1967 and in the data collected for the biological treatment studies.

The 12 composite samples collected in 1969 had an average phenol concentration of 36.4 mg/l. This level is supported by grab

TABLE 6

SAMPLING POINT C-7A

TOTAL COKE PLANT EFFLUENT

DATE	TIME TAKEN	TYPE OF SAMPLE	TEMP °C	pH	CONDUCTIVITY μ mhos/cm	AMMONIA NH_3 Mg/1	TSS Mg/1	TOC Mg/1
11-26-69	8:00am	Grab	40	8.5		217	510	
	9:00am	"	41	8.6		122	140	
	10:00am	"	40	8.8		169	95	
	11:00am	"	37	8.8		142	260	
	12:00pm	"	39	8.9	890	154	170	135.8
11-27-69	8:30am	"	36	8.7		200	1324	
	9:30am	"	44	8.3		81.4	110	
	10:30am	"	43	8.6		55.1	210	
	11:30am	"	40	8.4		34.2	292	
	12:30pm	"	34	8.9	800	171.8	518	
12-1-69	8:00am	"	41	8.5		133	-	
	9:00am	"	47	8.6		342	510	
	10:00am	"	36	8.4		382	1000	
	11:00am	"	35	8.4		346	350	
	12:00pm	"	49	8.8	1300	422	470	
12-10-69	8:00am	"	39	9.0		-	62	
	9:00am	"	39	9.0		-	41	
	10:00am	"	38	9.0		-	86	
	11:00am	"	38	8.9		-	59	
	12:00pm	"	45	9.0	950	261	88	
12-11-69	8:00am	"	39	8.8		-	66	
	9:00am	"	38	8.8		-	110	
	10:00am	"	36	8.6		-	110	
	11:00am	"	45	8.4		-	57	
	12:00pm	"	41	8.8		336	80	

TABLE 7

Plant Effluent, Sampling Pt. C-7A

Date	2-26-69	3-10-69	3-11-69	3-12-69	3-24-69	3-25-69	7-7-69	11-5-69
Time of Day	7:55 AM	1st sh	1st sh	1st sh	1st sh	1st sh	8:10 AM	1st sh
Type of Sample	Grab	Comp	Comp	Comp	Comp	Comp	Grab	Comp
Sampling Freq.	-	2 hrs	2 hrs	2 hrs	2 hrs	2 hrs	-	-
Temp., Range °C	29	28-40	38-41	33-42	33-43	31-41	41	-
pH	8.8	8.8	8.9	8.9	8.6	8.6	8.7	8.8
Cond., Range μ mhos/cm	-	-	-	-	1350-1650	1300-1700	-	-
P-Alk, $CaCO_3$ mg/1	100	100	140	210	110	--	294	180
M-Alk, $CaCO_3$ mg/1	324	360	400	502	510	446	536	406
Chloride, NaCl mg/1	30	80	60	110	60	118	24	--
Tot. Hard. $CaCO_3$ mg/1	166	102	150	132	150	160	-	178
Sulfate, SO_4 mg/1	-	-	-	-	-	-	96	375
Ammonia, NH_3 mg/1	99	139	-	-	-	-	-	138
Cyanide, CN^3mg/1	17	20	17	25	23	40	-	37
Phenol, C_6H_5OH mg/1	36	32	51	42	63	51	-	26
COD mg/1	688	621	307	602	1261	1198	611	-
Sulfide mg/1	3.5	5	1.5	9	22	7	-	-
Ether Extr. mg/1	31	56	31	5	19	20	-	-
TSS mg/1	2267	1056	395	617	680	198	145	150

Remarks: Ammonia still in operation by Nov. 5, 1969
Shift (sh)

TABLE 8

Sampling Point C-7A

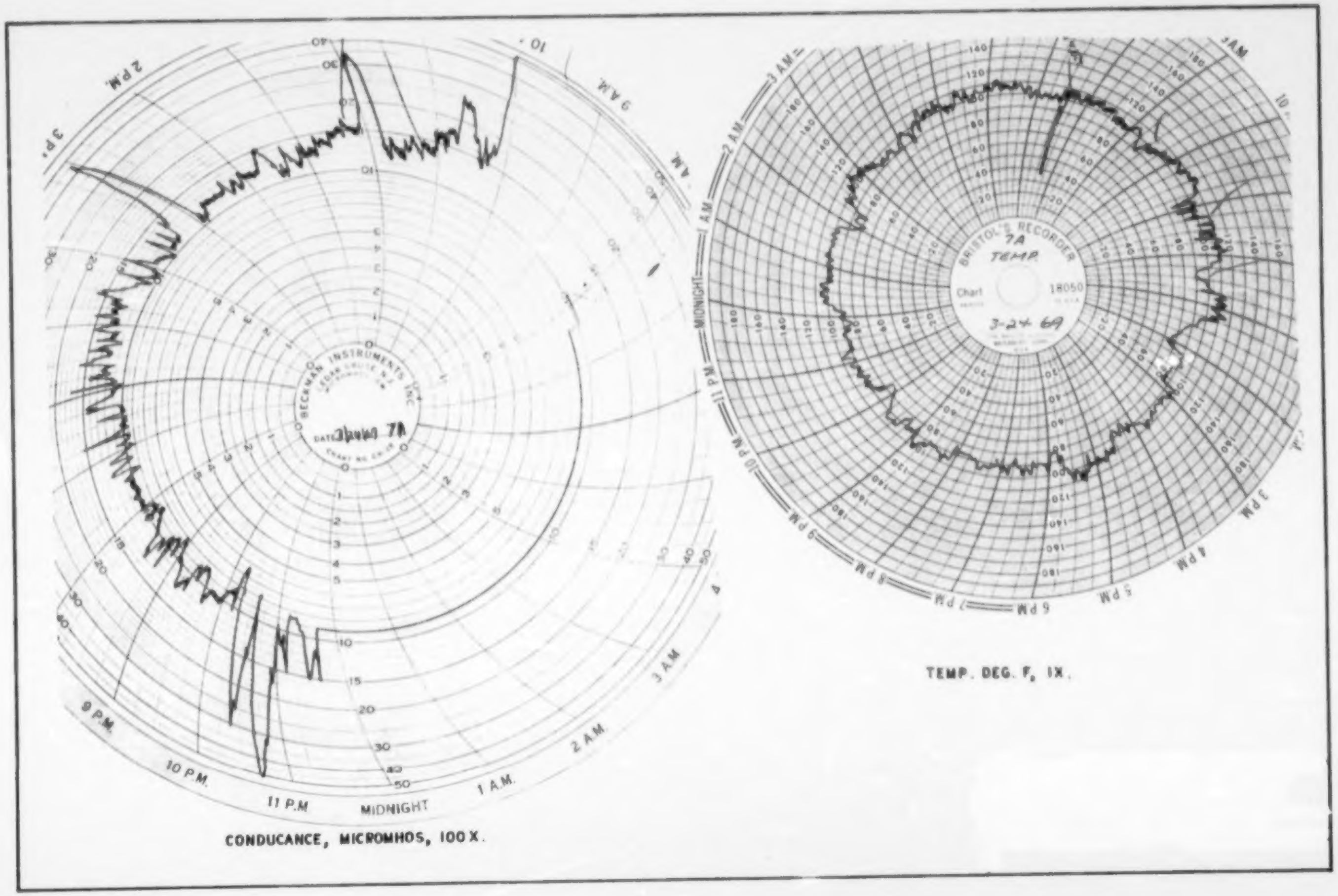
Total Coke Plant Effluent

Date	11-7-69	11-26-69	11-27-69	12-1-69	12-10-69	12-11-69
Time Taken	1st sh	8A - 12P	8:30A-12:30P	8A - 12P	8A - 12P	8A - 12P
Type of Sample	Comp	Comp	Comp	Comp	Comp	Comp
P-Alk, $CaCO_3$ mg/1	--	132	26	520	272	252
M-Alk, $CaCO_3$ mg/1	197	310	232	814	464	460
Chloride NaCl mg/1	146	66	52	122	136	160
Tot. Hard. $CaCO_3$ mg/1	188	80	100	109	86	91
Sulfate, SO_4 mg/1	379	230	206	242	267	670
O-Phosphate, PO_4 mg/1	--	0	0	2	0	0
Nitrate, NO_3 mg/1	--	0	0	0	--	--
TDS (Grav) mg/1	--	270	250	300	460	760
Cyanide, CN mg/1	29	8.5	8.1	14	22	22
Phenol, C_6H_5OH mg/1	11.5	28	19	27	41	31
Ether Extr. mg/1	--	11	19	12	12	12
TSS mg/1	1075	--	--	--	--	--
TOC mg/1	--	--	--	--	--	--

Heavy Metals
Analyses
3-12-1969

Mn: 0.1
Fe: 4.4
Cu: 0.01
Cr: <0.02
Ni: 0.02
V: <0.05
Zn: 1.3
Pb: <0.1

Remarks: Shift (sh)
Ammonia still operating



CONDUCANCE, MICROMHOS, 100 X.

TEMP. DEG. F, 1X.

Figure 5. TEMPERATURE AND CONDUCTANCE CHARTS - COMBINED COKE PLANT EFFLUENT SAMPLING POINT C-7A

sample data from 1964 to 1967; however, extremely low concentrations (<10 mg/l) and much higher concentrations (>100 mg/l) have been observed.

The 1969 data (Table 7) indicate considerable variation in COD.

Considerable variations in suspended solids and total solids have been observed. These are caused by discharge of spent wash water from the coal washery which are very heavy in suspended solids. The coal washery normally operates between 2:00 p.m. and 10:00 p.m.; however, some solids are discharged during maintenance and cleanup periods. Changes in suspended solids with the time of day is much in evidence.

Sample Point C-7B This is the total effluent from the benzol plant, the creosote plant, and the east side of the by-products plant. The data acquired at this point are presented in Table 9. There is considerable variation in COD. Composite samples from March, 1969 indicate low cyanide concentrations (2 - 9 mg/l) and phenol concentrations in the range 7 - 24 mg/l. High ammonia concentrations were the result of leakage of ammonium sulfate slurry from the saturator area. Oil and tar are evident at this point.

Sample Point C-8 The main discharge from the coal washery passes this point. It is essentially a heavy slurry of final refuse consisting of clay, silt, and some fine coal separated in the coal washing operation. The stream also contains up to 75 gpm from the pitch settling sump. Flows range from 600 - 750 gpm. TSS varies from 4% to 12% by weight.

Sample Point C-9 The flow at this point is intermittent. It normally contains high concentrations of suspended solids owing to drainage and overflow from the fine coal vacuum filter, refuse from the jig operation, and both fine coal and refuse from routine floor cleaning.

Sample Point C-11A This point receives 2 gpm from steam traps and 7 gpm from a coal sample washroom where zinc chloride is used. The sampling spot also receives 2 gpm from the discharge hose in a test oven installation. Steam traps also discharge in the oven area. Table 10 includes data collected.

Sample Point C-12 Water used in the Creosote plant is supplied by a re-cycle system. When a loss of power or pump failure occurs, the system automatically overflows to the sewer sampled at C-12. Data for this point, found in Table 2, reveal high concentrations of phenol.

Sample Point C-13 This station receives overflow from the creosote

TABLE 9

Sampling Point C-7B

Benzol Plant Effluent

Date	2-26-69	3-10-69	3-11-69	3-12-69	3-17-69	3-18-69	6-18-69	6-18-69	6-19-69
Time of Day	8:00 AM	1st sh	1st sh	1st sh	1st sh	1st sh	7:45 AM	1:30 PM	7:50 AM
Type of Sample	Grab	Comp	Comp	Comp	Comp	Comp	Grab	Grab	Grab
Sampling Freq.	--	2 hrs	2 hrs	2 hrs	2 hrs	2 hrs	--	--	--
Temp., Range ^{o}C	14	19-30	75-31	26-27	23-27	25-31	22-31	29	30.0
pH	8.4	8.7	8.4	8.7	7.5	8.5	7.4	7.7	7.4
Cond. μ mhos/cm	--	--	--	--	1000	--	--	1050	--
Min-Max	--	--	--	--	600-5000	--	--	600-4200	--
P-Alk, $CaCO_3$	10	50	40	28	--	40	0	0	0
M-Alk, $CaCO_3$	128	200	240	258	184	220	84	112	120
Chloride, NaCl mg/l	20	10	40	40	10	40	35	14	40
Tot.Hard. $CaCO_3$ mg/l	194	144	150	164	196	132	116	110	160
Sulfate mg/l	--	--	--	--	--	--	--	--	152
TDS (grav) mg/l	--	--	--	--	--	--	387	599	454
Ammonia, NH_3 mg/l	19	79	198	343	331	98	--	--	--
Cyanide, CN^- mg/l	2	9	5	5	3	3	--	--	--
Phenol, C_6H_5OH mg/l	7	20	22	21	12	24	--	--	--
COD mg/l	160	335	1075	345	727	527	--	119	--
Sulfide mg/l	<0.5	1.5	5.0	3.5	9.4	1.5	--	--	--
Ether Extr. mg/l	41	95	33	5	13	17	--	--	--
TSS mg/l	36	197	138	347	40	124	79	52	49
BOD mg/l	--	--	--	--	--	--	--	96	240

Remarks: No caustic dumps in these samples

TABLE 10

Sampling Points C-11A and C-13

Source of Sample	C-11A	C-11A	C-13	C-13	C-13	C-13	C-13	C-13
Date	6-9-69	6-9-69	2-24-69	2-28-69	3-3-69	3-27-69	5-13-69	6-9-69
Time of Day	8:10 AM	8:10 AM	8:00 AM	8:00 AM	8:10 AM	1st sh	8:00 AM	8:00 AM
Type of Sample	Grab*	Grab**	Grab	Grab	Grab	Comp	Grab	Grab
Temp., Range °C			43	40.5	40.5	35-46		
pH	5.5	7.7	9.0	8.8	9.3	9.0	8.7	--
P-Alk, $CaCO_3$ mg/l	0	0	--	--	1850	1410	--	--
M-Alk, $CaCO_3$ mg/l	10	20	--	--	4260	3240	--	--
Chloride, NaCl mg/l	--	10	--	--	90	--	--	--
Tot Hard., $CaCO_3$ mg/l	--	--	--	--	60	70	--	--
Ammonia, NH_3 mg/l	--	--	2006	1921	1972	1258	1113	1788
Cyanide, CN^- mg/l			92	86	71	37	9.5	20
Phenol, C_6H_5OH mg/l			571	561	1300	376	665	590
COD mg/l	286	--	--	--	4470	3269	--	--
Sulfide mg/l	--	--	--	--	--	30	--	--
Ether Extr. mg/l	--	--	--	--	7590	--	--	--
TSS mg/l	815	152	--	--	1500	219	--	--

Remarks: * Coal Sampler
** Water Heater

re-cycle sump when the re-circulating pump is not working. It receives excess ammonia flushing liquor when the ammonia still is not operating. Table 10 shows the heavy contamination present in this stream.

Sample Point C-15 This point receives seal water from the gas holder. It is possible that the stream contains water soluble components of the gas. The flow is normally 1-2 gpm. Occasionally 75 gpm from the primary cooler system used for valve freezing protection enters the sewer at this location.

As shown in Table 14, the gas holder sealing water can become contaminated with ammonia, cyanide and phenols. However, the degree of contamination is not in excess of that found in the combined coke plant effluent, sampling point C-7A.

Sample Point C-16 This is an overflow connection for the primary cooler which handles primary cooler water in excess of the sewer capacity at the wire mill. Water from a 3/4" seal on the cleaned gas returning to the coke batteries flows into the sewer at C-16 at a rate of 5 gpm. There is also a 17 gpm component from "E" battery (gas seals and condensates). Both small flows show slight contamination in Table 14.

Sample Point C-18 This point receives approximately 100 gpm of uncontaminated water from steam traps and seals on steam turbine pumps. There is a fire station in this area having standby pumps. Seal water is discharged from these pumps continuously.

Sample Point C-18A This point represents the combined discharge from the coal washery, the north quench, and the coke plant area east of the By-Products Building. Total flow may reach 2300 gpm. Tables 11-13 contain pertinent data.

C-18A contains high concentrations of ammonia, cyanide, phenol, and COD. This is to be expected inasmuch as the sampling point is downstream from points C-1, 2, 3, 8, 11, 13, 14, 15, and 16. The source of ammonia is ammonia flushing liquor. Phenol, ammonia and COD concentrations at this point are generally higher than those experienced in the combined coke plant effluent (C-7A).

Sample Point C-28 The stream flow at this point is composed of seal water from the exhauster in the by-products plant (47 gpm), air compressor water from the by-products plant(17 gpm), and condensate from a nearby condenser for exhaust steam (33 gpm). It is expected that the seal water contains impurities from contact with the gas. Table 14 reveals the relative purity of the waste stream at this point.

TABLE 11

Sampling Point C-18A

Date	3-17-69	3-18-69	3-23-69	3-24-69	3-25-69	6-19-69	7-7-69	11-5-69
Time of Day	1st sh	1st sh	1st sh	1st sh	1st sh	8:05 AM*	9:10 AM	1st sh
Type of Sample	Comp	Comp	Comp	Comp	Comp	Grab	Grab	Comp
Sampling Freq.	2 hrs	2 hrs	2 hrs	2 hrs	2 hrs	--	--	--
Temp., Range ^{o}C	26-39	34-50	52-53	34-52	39-53	30	46	--
pH	8.9	9.0	9.0	8.8	8.6	9.0	9.0	8.7
Cond. µ mhos/cm	--	550-3200	--	--	--	--	--	--
P-Alk, $CaCO_3$ mg/1	326	360	360	200	198	400	528	--
M-Alk, $CaCO_3$ mg/1	816	890	882	970	694	960	1158	418
Chloride, NaCl mg/1	120	160	120	160	120	--	190	160
Tot. Hard. NaCl mg/1	88	126	70	90	88	150	56	182
Sulfate, SO_4 mg/1	--	--	--	--	--	--	71	--
Nitrite, mg/1	--	--	--	--	--	--	.8	--
TDS (Grav) mg/1	--	--	--	--	--	305	1224	--
Ammonia, NH_3 mg/1	378	391	408	612	--	--	--	179
Cyanide, CN mg/1	8	18	15	26	42	--	--	55.5
Phenol, C_6H_5OH mg/1	116	130	147	194	136	--	--	50
COD mg/1	1735	1344	927	1734	890	--	2240	--
Sulfide mg/1	8	9	10	19	9	---	--	--
Ether Extr. mg/1	19	25	21	26	27.5	--	--	--
TSS mg/1	1235	2167	663	103	1562	558	460	250

Remarks: Ammonia still in operation by Nov. 5, 1969
*Final Cooler bypassed to sewer

TABLE 12

Sampling Point C-18A

All Flows Except South Quench & Benzol Yard

Date	11-6-69	11-7-69	11-26-69	11-27-69	12-1-69	12-10-69	12-11-69
Time Taken	1st sh	1st sh	8A - 12P	8:30A-12:30P	8A -12P	8A - 12P	8A - 12P
Type of Sample	Comp	Comp	Comp	Comp	Comp	Comp	Comp
pH	8.1	8.6	--	--	--	--	--
P-Alk, $CaCO_3$ mg/l	--	--	226	30	508	284	232
M-Alk, $CaCO_3$ mg/l	--	333	450	206	780	536	452
Chloride, NaCl mg/l	--	47	82	36	112	120	88
Tot. Hard., $CaCO_3$ mg/l	--	146	72	92	62	69	71
Sulfate, SO_4 mg/l	--	--	178	148	187	170	178
O-Phosphate, PO_4 mg/l	--	--	0	0	0	0	0
Nitrate, NO_3 mg/l	--	--	0	0	0	--	0
TDS (Grav) mg/l	--	--	230	230	250	270	250
Cyanide, CN mg/l	85	81	13	32	8.6	15	14
Phenol, C_6H_5OH mg/l	32	31	49	27	46	74	56
Ether Extr. mg/l	--	--	11	25	10	11	7
COD, mg/l	406	--	--	--	--	--	--
TSS, mg/l	208	4597	--	--	--	--	--
TOC, mg/l	178	--	--	--	--	--	--

TABLE 13

Sampling Point C-18A

All Flows Except South Quench and Benzol Yard

Date	Time Taken	Type of Sample	Temp °C	pH	Conductivity µ mhos/cm	Ammonia NH mg/1	TSS mg/1	TOC mg/1
11-26-69	8:00 am	Grab	54	8.9	--	73	280	--
	9:00 am	Grab	54	8.6	--	154	120	--
	10:00 am	Grab	55	8.9	--	196	470	--
	11:00 am	Grab	54	8.8	--	212	220	--
	12:00 pm	Grab	49	8.8	1095	216	130	182
11-27-69	8:30 am	Grab	26	8.6	--	150	--	--
	9:30 am	Grab	44	8.4	--	30	--	--
	10:30 am	Grab	45	8.7	--	46	177	--
	11:30 am	Grab	38	8.7	--	83	320	--
	12:30 pm	Grab	52	8.9	725	139	321	104
12-1-69	8:00 am	Grab	47	8.5	--	108	--	--
	9:00 am	Grab	47	8.6	--	427	220	--
	10:00 am	Grab	50	8.8	--	427	330	--
	11:00 am	Grab	52	8.8	--	311	230	--
	12:00 pm	Grab	47	8.9	1350	1525	250	--
12-10-69	8:00 am	Grab	58	9.1	--	--	82	--
	9:00 am	Grab	58	9.1	--	--	100	--
	10:00 am	Grab	55	9.1	--	--	120	--
	11:00 am	Grab	59	9.0	--	--	110	--
	12:00 pm	Grab	57	9.0	950	249	120	--
12-11-69	8:00 am	Grab						
	9:00 am	Grab						
	10:00 am	Grab						
	11:00 am	Grab						
	12:00 pm	Grab						

TABLE 14

Sampling Points C-15, C-16, C-28

	Gas Holder		Primary Cooler Overflow	North West By-Product Building			
Sample Source	C-15	C-15	C-16	C-28	C-28	C-28	C-28
Date	3-27-69	7-7-69	6-19-69	6-18-69	6-19-69	2-27-69	3-23-69
Time of Day	1st sh	9:15 AM	8:10 AM	1:20 PM	7:20 PM	8:00 AM	7A - 12P
Type of Sample	Comp	Grab	Grab	Grab	Grab	Grab	Comp
Temp Range °C	51-56	52	18	27	28	23	9-10
pH	8.1-8.9	8.2	7.9	7.6	7.6	8.3	8.1
P-Alk, $CaCO_3$ mg/1	100	6	0	0	0	10	0
M-Alk, $CaCO_3$ mg/1	280	100	86	94	110	102	128
Chloride, NaCl mg/1	20	18	40	20	30	60	24
Tot. Hard., $CaCO_3$ mg/1	120	88	160	136	160	172	208
TDS (Grav) mg/1	--	129	568	220	358	--	--
Ammonia, NH_3 mg/1	85	--	--	--	--	7	13
Cyanide, CN^3 mg/1	5	--	0.3	--	--	0.3	0.1
Phenol, C_6H_5OH mg/1	21	--	.07	--	--	.20	.25
Sulfide mg/1	--	--	--	0.1	0.1	--	--
COD, mg/1	385	214	--	--	--	95	161
Ether Extr., mg/1	--	--	--	--	--	135	51
TSS, mg/1	396	397	47	31	39	26	96

Sample Point C-28A This sampling point normally consists of water from the interceptor sump in the creosote plant (C-5). Uncontaminated overflow from the wash oil coolers and some primary cooler water can be diverted from the Wire Mill Pond area through movable gates near the wash oil coolers. Analyses of flow from the creosote interceptor sump (C-5) describe point C-28A and are found in Table 2.

Sample Point C-29A Uncontaminated water from steam traps (26 gpm) and cooling water from roof condensers in the new Benzol Building (744 gpm) enter the sewer at this point. An additional 114 gpm from the light oil separator reports to this manhole and introduces significant concentrations of COD. Data on these flows are compiled in Tables 15 and 16.

Sample Point C-30 This point receives discharge from the saturator and fertilizer plant areas. Drainage from this area could be highly contaminated because of excess ammonia slurry and other concentrated chemicals used in the process including acid. In addition C-30 receives the discharge from the agitator area where caustic is used for neutralization. C-30 also receives accidental spills from tank loading operations in the immediate vicinity. These tank cars carry acid, naphthalene, and caustic. The acid sludge from the agitators is removed in the tank car.

The flow at C-30 is 25 gpm when the fertilizer plant is not operating. This particular flow originates in water seals on valves and gas lines and from water hoses. Pertinent data appear in Table 17.

Sample Point C-31 This flow contains caustic and oil from the neutralization of light oil following the acid wash. Normally 250 gallons per day is discharged on a batch basis.

Sample Point C-32 Flow at this point contains water from water hoses and water seals on gas lines. In addition drainage from the Fertilizer Plant and Saturator areas flow through this manhole. The stream characterisitics at this point are best represented by C-30 data found in Table 17.

Sample Point C-33 This point receives flows from the coal washery, the north and south quench and the east portion of the by-products building. There is also a 25 gpm gas drip pot discharge to this point which is characterized by data presented in Table 16.

Miscellaneous Tests

In addition to the data mentioned above, special analyses for IDOD were made at selected points of sampling. Data for IDOD and sulfite are presented in Table 18. This data indicates that the waste at some of the sampling points exerts an immediate oxygen demand on a system designed for biological oxidation of

TABLE 15

Sampling Points Misc.

Source of Sample	Wash Oil Pump Seal		Exhauster Relief Seal		By-Prod. Bldg. Combination		Condenser New Benzol Bldg.	
Date	6-18-69	6-19-69	6-18-69	6-19-69	6-19-69	7-7-69	6-18-69	6-18-69
Time of Day	1:25 PM	7:25 AM	1:30 PM	7:30 AM	7:35 AM	8:00 AM	1:00 PM	7:37 AM
Type of Sample	Grab	Grab	Grab	Grab	Grab	Grab	Grab	Grab
Temp. ^{o}C	30	30	27	27	28	30	--	--
pH	7.6	7.7	7.6	7.6	7.4	7.8	7.6	7.6
P-Alk, $CaCO_3$ mg/1	0	0	0	0	0	0	0	0
M-Alk, $CaCO_3$ mg/1	92	100	92	90	86	90	84	94
Chloride, NaCl mg/1	12	14	10	14	12	12	12	14
Tot. Hard., $CaCO_3$ mg/1	136	150	136	150	160	81	140	136
Nitrates, NO_2 mg/1	--	--	--	--	--	1.4	--	--
TDS (Grav) mg/1	255	249	190	305	235	224	290	244
Ammonia, NH_3 mg/1	--	--	--	--	4.3	--	--	--
Cyanide, CN mg/1	--	--	--	--	.02	--	--	--
Phenol, C_6H_5OH mg/1	--	--	--	--	.05	--	--	--
COD, mg/1	--	--	--	--	--	16	--	--
TSS, mg/1	30	25	54	64	73	134	59	53
BOD, mg/1	--	--	--	--	10.5	--	--	--

TABLE 16

SAMPLING POINTS C-29A, Misc

	Light Oil Separator				Roof Condensor New Benzol Building			C-33 Drip Pot East "E" Battery
Source of Sample	C-29A	C-29A	C-29A	C-29A				
Date	4-9-69	6-18-69	6-18-69	7-7-69	6-18-69	6-18-69	7-7-69	7-7-69
Time of Day	8:00 AM	7:40 AM	1:15 PM	8:15 AM	7:30 AM	1:10 PM	8:20 AM	9:00 AM
Type of Sample	Grab	Grab	Grab	Grab	Grab	Grab	Grab	Grab
Temp ^{o}C	34	--	--	44	--	--	30	26
pH	6.8	8.1	7.5	7.6	7.6	7.7	7.7	7.2
P-Alk, $CaCO_3$ mg/l	0	18	0	0	0	0	0	0
M-Alk, $CaCO_3$ mg/l	14	358	238	144	82	98	100	62
Chloride, NaCL mg/l	10	12	12	40	5	6	12	10
Tot. Hard., $CaCO_3$ mg/l	20	100	100	76	138	134	132	--
Sulfate, SO_4 mg/l	--	--	--	40	--	--	54	49
Nitrite, NO_3 mg/l	--	--	--	2.6	--	--	1.9	.8
TDS (Grav) mg/l	--	240	197	194	377	254	187	144
Ammonia, NH_3 mg/l	11	--	--	--	--	--	--	--
Cyanide, CN mg/l	1.4	--	--	--	--	--	<0.1	.55
Phenol, $C6H_5OH$ mg/l	0.4	--	--	--	--	--	<0.1	.10
COD, mg/l	1065	--	--	659	--	--	23	--
Sulfide, mg/l	1	--	--	--	--	--	--	--
Ether Extr. mg/l	7.4	--	--	--	--	--	--	--
TSS, mg/l	94	33	27	33	49	35	99	14
BOD, mg/l	--	--	--	--	--	--	44	--

TABLE 17

Sampling Points C-30 and C-32

Ammonia Saturators and Fertilizer Plant Drainage

Source of Sample	C-30	C-30	C-30	C-30	C-30	C-30	C-32	C-32	C-32
Date	2-27-69	2-27-69	4-9-69	6-18-69	6-19-69	7-7-69	6-9-69	6-18-69	7-7-69
Time of Day	8:10 AM	10:00AM	8:00 AM	7:30 AM	7:35 AM	8:25 AM	8:00 AM	7:15 AM	9:10 AM
Type of Sample	Grab	Grab	Grab	Grab	Grab	Grab	Grab	Grab	Grab
Temp °C	40	41	43	42	46	43	--	--	54
pH	6.2	7.7	7.5	5.7	7.0	2.7	7.9	8.5	7.8
P-Alk, $CaCO_3$ mg/l	0	0	0	0	0	0	0	22	0
M-Alk, $CaCO_3$ mg/l	14	158	100	--	114	--	40	158	84
Chlorides, NaCl mg/l	40	40	30	60	50	--	50	12	9
Tot. Hard. $CaCO_3$ mg/l	--	190	190	--	140	156	--	112	108
Sulfates SO_4 mg/l	--	--	--	--	--	--	--	--	35
Nitrites NO_3 mg/l	--	--	--	--	--	--	--	--	.8
TDS (Grav) mg/l	--	--	--	2813	6488	4157	--	336	191
Ammonia, NH_3 mg/l	1088	--	782	--	1630	--	--	24	16
Cyanides, CN mg/l	.3	--	2	--	1.5	--	--	.7	.4
Phenol, C_6H_5OH mg/l	4.75	--	3.2	--	2.5	--	--	0.1	0.1
COD mg/l	600	2270	1860	--	--	1770	311	48	116
Ether Extr. mg/l	20	--	40	--	--	--	--	--	--
TSS mg/l	800	720	154	232	155	48	199	56	212
BOD mg/l	--	--	--	--	880	--	--	--	--

TABLE 18

Summary of Immediate Oxygen Demand

Sulfite Values, misc. sampling points
September 23, 1969

Stream	Point	IDOD	SO_3
1. IDOD Values			
Light Oil Separator	C-16	.6 mg/l	-
Creosote Plant	C-5	2.5 mg/l	-
	C-30	28.4 mg/l	-
	18-A	6.4 mg/l	-
	C-3	3.9 mg/l	-
2. Sulfite Values			
Combined Effluent	C-7B	-	1.5 mg/l
	C-30	-	2.5 mg/l
Light Oil Separator	C-29A	-	35 mg/l
Roof Condenser	C-29A	-	tr
Top Condenser	C-29A	-	tr
Condensate	C-28	-	tr
By-products	C-28	-	tr

pollutants such as phenols. The initial oxygen demand could be caused by inorganic chemical reducing agents such as sulfur dioxide, ferrous iron, sulfide, or sulfite.

Discussion:

At the present time as can be seen from the data in Table 1 the flow rate of the combined coke plant effluent is highly variable. It may vary from 1700 gpm to 3400 gpm. The variation in flow rate is largely due to intermittent operations such as the coal washery and the manual control of water flows. Data collected during this survey and the prior flow data accumulated by CF&I indicate that the average normal flow rate is 3000 gpm.

Concentrations of the principal chemical pollutants also vary widely apparently due to both flow variations and operating changes. Because of these variations it is difficult to determine accurately the overall loadings. However, the data obtained from the survey and the pilot testing may be used to estimate the total effluent concentrations.

The flow at point C-7A is approximately 3000 gpm, assuming that this point includes discharge of excess flushing liquor without operation of the ammonia stills but does not include the flow from the coal washery. On this basis a normal plant concentration of 25-35 mg/l phenol, 15 to 25 mg/l CN, and 180-250 mg/l NH_3 can be predicted. This amounts to an average of 721 Lbs/Day CN, 1080 Lbs/Day phenol and 7710 Lbs/Day of NH_3. Variations may be as much as plus or minus 50%. These variations could be eliminated through the use of an equalization basin.

The survey pinpointed all existing and potential sources of pollution and it also determined several clean water streams which may be diverted into existing clean sewers and thus reduce the final treatment volume. Table 19 presents a summary of these streams as well as those which may be eliminated by process change.

Diversion of uncontaminated streams from the future final treatment plant would result in a flow rate of approximately 1200 gpm at point C-7A. Total contaminant loadings would remain essentially the same while the concentrations would increase by a factor of 3. Although the flow variations may be reduced by operational changes they will still exist. The recommended operation of the ammonia stills will add an additional 100-150 gpm flow to point C-7A. It is therefore the recommendation of this study that the final treatment plant be designed for a flow of 2000 gpm. Before finalization of the treatment plant design the recommended piping and process changes should be completed and actual flow and concentration data be reviewed.

TABLE 19

SUMMARY OF STREAMS TO BE DIVERTED FROM COMBINED PLANT EFFLUENT

Clean Streams to be Diverted

Sample Point	Source	Avg Flow gpm	Temp °F
C-1	Individual Oil Cooler	25	70
C-18	Steam Turbine Condensate	100	130
C-28	Steam Condensate	33	130
C-29A	Benzol Plant Indirect Cooling	744	115
C-29A	Steam Condensate	26	180
	TOTAL	928	

Streams to be Eliminated by Process Changes

Sample Point	Source	Method of Elimination	Hour gpm
C-8	Coal Washery	Local Treatment	600
C-29A	Light Oil Separator	Use for Final Cooler	114
C-3 & C-4	Coke Quencher	Elimination of Excess Use	
C-30	Caustic Dump	Recycle	-
		TOTAL	714

TREATABILITY STUDY

BIOLOGICAL OXIDATION

Activated Sludge

The feasibility of treating the combined effluent from the coke plant (C-7A) in a biological treatment system was investigated in a pilot plant operated adjacent to that sampling point. The pilot unit was designed and installed in accordance with the data contained in the Appendices D & E, pp. 252-257. Initial tests were run using an activated sludge system operating under conditions of plug flow.

Since the pH of the waste water from sampling point C-7A ranged from 7.5 to 9.0, phosphoric acid was used to reduce the pH to the range 6.5-7.5. The phosphoric acid also provided the phosphorus nutrient requirement necessary for the system to function. Nitrogen was supplied by the NH_3 present. A minimum ratio of BOD (100 mg/l): Nitrogen (5 mg/l): Phosphorus (1 mg/l) was maintained.

The pilot unit is illustrated in Figures 6 and 7. Primary settling occurred in the feed reservoir located next to the sampling point. The reservoir was filled with waste from C-7A at a time when the coal washery was not in operation and when no caustic or acid dumps were sent to the sewer in the benzol plant. A feed pump transferred the liquid from the reservoir to the first stage of the activated sludge unit - the primary clarifier (capacity 8 gallons).

The mixed liquor was contained in a 45 gallon tank divided into six equal volume compartments, each equipped with an aeration device. Flow from one compartment to another was facilitated through openings in the dividing walls. The openings were located alternately near the bottom and near the top. Overflow from the final compartment passed to the secondary clarifier (capacity 8 gallons).

The secondary clarifier was originally designed with an air lift to continuously transfer settled sludge to the No. 1 aeration chamber. A small centrifugal pump operated by a two minute timer was substituted for the air lift soon after initial start up.

Initially the solids separation in the secondary clarifier was inadequate. Additional retention time was provided by using 55 gallon drums filled in parallel during a normal day of operation. After each 8 hours of operation, the drums were decanted and the sludge collected from the drums was returned to the aeration unit.

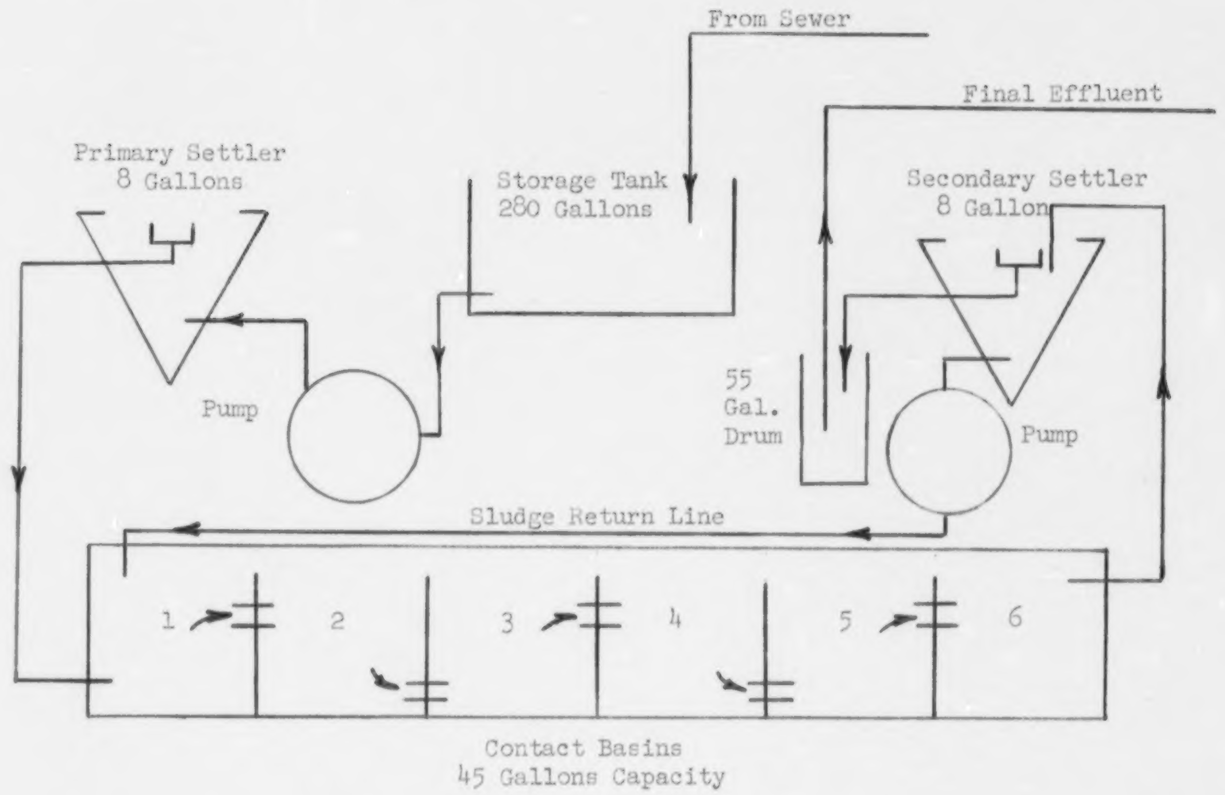

From Sewer
Final Effluent
Primary Settler
8 Gallons
Storage Tank
280 Gallons
Secondary Settler
8 Gallon
55
Gal.
Drum
Pump
Pump
Sludge Return Line
1
2
3
4
5
6
Contact Basins
45 Gallons Capacity

Figure 6. TRAILER MOUNTED ACTIVATED SLUDGE SYSTEM

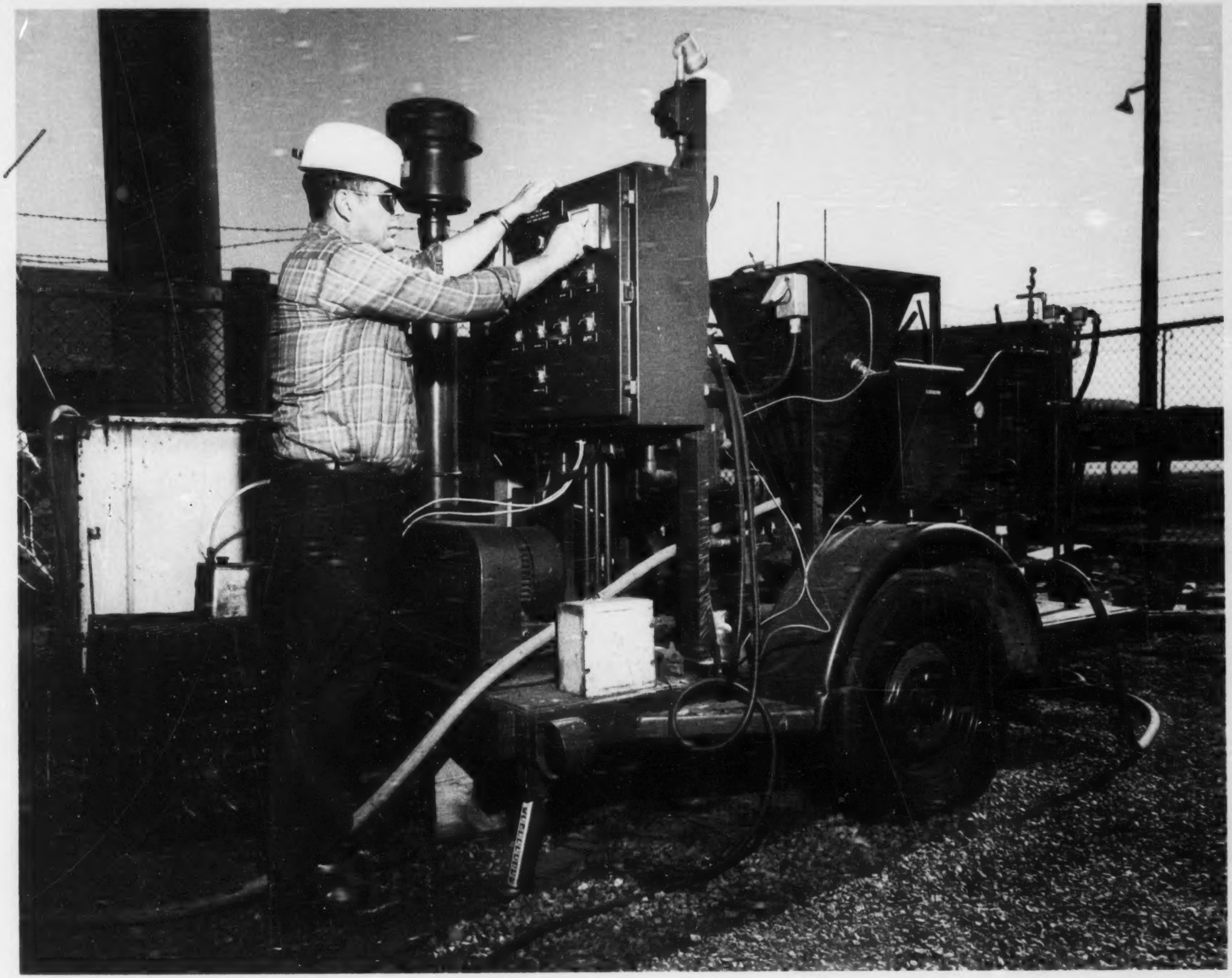


Figure 7. BIOLOGICAL TRAILER MOUNTED PILOT PLANT

Flow Measurement

Flow measurements were made by timing the volume in a graduated beaker at two points - the feed line from the primary settler and the sludge return line. The flows were intermittent because the influent flow was controlled by a timer and solenoid valve while the sludge return was controlled by a timer operating a small centrifugal pump.

Sampling Program

Grab and composite samples were taken from influent and effluent flows. Normally grab samples were collected every morning and analyzed for pH, dissolved oxygen, mixed liquor suspended solids, volatile suspended solids, settleable solids, phenol, cyanide, COD, and ammonia nitrogen. Detention time (determined from the flow into the aeration basins), efficiency, and pounds of phenol, cyanide, and COD removed were calculated from the analytical data. Additional spot checks were made on phosphate concentrations. Temperatures were recorded on a 24-hour basis.

Acclimation of Sludge

During the course of the biological studies, it became necessary on several occasions to obtain sludge for acclimation to the Coke Plant wastes. Sludge was obtained from the Pueblo Municipal Sewage Treatment Plant's trickling filter operation, from CF&I's modified extended aeration sanitary sewage treatment system, and from the area where CF&I's effluent stream enters the Arkansas River. Five gallons of sludge was mixed with 20 gallons of raw municipal sewage in a 55 gallon drum equipped with an air diffuser. A 50 ml sample of ammonia flushing liquor was added to the barrel and aeration was continued for 24 hours.

After 24 hours, the 55 gallon drum was filled with plant water to within 15 inches of the top, at which time 150 ml of ammonia flushing liquor and 5 grams of dextrose were added. The pH was adjusted to 7.0 with phosphoric acid and aeration was continued. After the second 24 hour aeration period, a daily procedure was established for developing the acclimated culture. Aeration was stopped and the sludge was settled for 1 hour. The top 50-60% of the liquid was siphoned away and the drum was refilled with mill water. The increments of ammonia flushing liquor were gradually increased over a ten day period. The pH was maintained at 7.0 with phosphoric acid. It was necessary to aerate to reduce the ammonia concentration below 2000 mg/l, to cool the ammonia flushing liquor, and to remove tar from the feed. The acclimation schedule is presented on page 52.

Day	ml of NH_3 Liquor	Grams of Dextrose
1	50	0
2	150	5
3	2000	40
4	3000	40
5	4500	30
6	6000	30
7	7500	20
8	9000	10
9	10,000	10
10	10,000	0

Acclimation was followed by phenol reduction. The initial concentration of phenol was measured each day after the ammonia liquor was added and mixed, and prior to siphoning.

Plug Flow Activated Sludge

Initially difficulties were encountered in developing an activated sludge colony. The pilot unit was operated continuously as a very low flow rate, starting at approximately 17 hours detention time. Results from this operation are shown on Table 20, Page 55.

The first test series was run from April 21, to April 24, 1969. During this time the sludge volume (mixed liquor settleable solids) amounted to approximately 16 ml/l. Mixed liquor volatile suspended solids (MLVSS) was also low. The phenol reduction was in the neighborhood of 99%, although it dropped at the end of the series. The inlet cyanide concentration stayed below 4 mg/l during the first three days. A reading of 7.2 mg/l cyanide was taken on the last day.

Another test series was conducted between April 28 and May 1, 1969. The retention time was still high; the flow rate indicated 14.2 hours detention. The settleable solids concentration was somewhat lower than before. Phenol reduction was 98% and cyanide reduction was 99%.

The next series was conducted from May 19-22. The detention time was reduced to approximately 10 hours. The sludge volume was very low. Readings of 2.5 mg/l settleable solids indicated essentially no settleable organic material present. In spite of this, phenol and cyanide reductions were good.

Phenol, cyanide, and COD removals expressed as pounds removed per pound of mixed liquor volatile suspended solids per day are very low and are included in Table 20, p 55.

Another series was run May 27-29, again with very little sludge production.

During all the test studies, the dissolved oxygen in the mixed liquor tanks was maintained at 6-7 mg/l. It became apparent that the poor settling characteristics of mixed liquor suspended solids was the result of excessive aeration which caused floc dispersion to such a degree that good settling was not possible. The aeration rate was decreased to a level which produced a dissolved oxygen concentration of 2 mg/l or less. However, difficulties were encountered in keeping the sludge in suspension. Re-adjustment of the oxygen level to approximately 5 mg/l prevented the sludge from settling out in the aeration tank.

At this rate of aeration the settleable solids improved as seen in the data taken June 5-11. The detention time was reduced to approximately 7 hours. Phenol removal remained excellent while cyanide removal was 66%.

Additional values for phenol and cyanide loadings were obtained, as shown in the table. The sludge volume, during the bulk of the test, varied from 18-50 ml/l.

Mechanical failure of the sludge return system took place at this time causing a large drop in sludge volume in the aeration tanks.

After the sludge had been redeveloped, another test series was started on July 7. Detention time was reduced to 5 hours. Mechanical difficulties with the feed system resulted in flow variation. Phenol and cyanide reduction remained excellent during the higher flow rate tests. Phenol removal of .19 and .35 Lbs/Lb MLVSS/Day were obtained. Cyanide removals expressed in the same terms were less than 0.4. Sludge volumes remained high (30-45 ml/l). During this series the dissolved oxygen content was reduced to the range 1.6-4.6 mg/l. The sludge had good settling characteristics and was brown in color which by now had been found to be typical of high activity.

Another series was started on July 14 and included more frequent analyses. The detention time was maintained at 5 hours. The initial phenol concentration was approximately 50 mg/l, and the cyanide concentration remained fairly constant at approximately 4 mg/l.

A gradual dropoff of influent phenol concentration was noted as shown in the table. The reduced phenol concentration in the influent resulted in a corresponding decrease in phenol loading as would be expected. The cyanide loading stayed constant since there was essentially no reduction in the cyanide concentration in the feed.

In spite of the reduction of phenol in the influent, the sludge concentration and volume remained high although there was a decrease at the end of the series. At this time the feed tank had apparently become contaminated with bacteria which consumed the phenol before the solution was put through the pilot unit.

Between July 17 and September 5 the system was operated continuously with a gradual decrease in detention time. Phenol removal in excess of 99% was accomplished with an influent phenol range of 1-42 mg/l. The sludge volume remained above 40 ml/l and exhibited excellent settling characteristics. The phenol loading varied considerably because of the fluctuation in influent concentrations. However, values between .09 and .16 Lbs Phenol/Lb MLVSS/Day were common.

A series of loading tests were started on August 18. The objective of this study was to determine phenol removal as a function of MLVSS and detention time. Results are shown in Table 22.

The plant was equipped with a feed reservoir having a capacity equivalent to approximately 80 hours of operation at maximum detention time (4.15 hrs). Phenol removal was calculated directly from the feed concentration and the total amount of phenol found in the effluent.

Detention time was decreased stepwise and the operation was continued at each fixed flow rate for at least two days to insure stabilization of the system. While the MLVSS varied considerably with flow rate, variation of solids at a given flow rate was not significant.

As shown in Table 22, a detention time of 1.94 hours on September 5, resulted in reduced phenol removal and loss of sludge. At this time phenol removal was up to .4 Lbs/Lb MLVSS/Day. The high figure resulted from a reasonably steady feed of 30-39 mg/l of phenol.

In all of the above tests it became evident that one of the major problems in maintaining a biological system was the low concentration of phenol in the influent. Phenol is the main source of carbon for the microorganisms and even in situations when phenol and cyanide concentrations are nearly equal, phenols contain 77% carbon as compared to cyanide 46% carbon. Reduction in initial phenol concentration would greatly hamper biological activity.

No attempt was made to control the temperatures in the pilot system. Waste water temperatures at the time samples were pumped from the sewer were between 36-40 °C. From this point, temperatures gradually decreased until a new sample was obtained. During the months of April and May system temperatures at night were as low as 7 °C and ranged from 10-21 °C during the day.

Table 20

Results of Activated Sludge Pilot Plant Operation

Date	Detention Time, Hrs.	Influent NH₃ mg/l	% NH₃ Red.	Infl. Phenol mg/l	% Phenol Red.	Influent CN mg/l	% CN Red.	% COD Red.	Settleable Sol. ml/l	System Solids MLSS LBS.	MLVSS mg/l	lbs.	Removal LBS/LB MLVSS/Day Phenol	Cyanide	COD
4-21	17.7	91.5	36.6	17.5	99+	3.6	39	-	16	.19	-		-	-	-
4-22	17.7	85	35	16.5	97	2.0	15	-	16	-	-		-	-	-
4-23	17.7	89.3	5	8.5	99	3.0	43.3	-	10	-	-		-	-	-
4-24	14.2	225	7.0	29.4	73	7.2	71	-	14	.18	-		-	-	-
4-28	14.2	238	7.2	28.4	98	6.16	99	-	5	.05	-		-	-	-
4-29	14.2	229	7.1	37.6	-	10.2	-	-	10	-	-		-	-	-
4-30	14.2	229	0	33.3	-	5.08	-	-	15	-	-		-	-	-
5-1	14.2	225	1.5	9.5	98	6.06	99	-	9	.16	-		-	-	-
5-19	10.3	78	0.0	4.8	98	3.72	78	17	5	.43	239	.09	.04	.03	1.64
5-20	10.3	105	45.7	2.7	96	2.0	85	24	2.5	.19	189	.07	.03	.02	2.3
5-21	9.7	82	6.1	30	87	5.97	78	65	2.5	.13	116	.06	.42	.07	.05
5-22	10.6	-	-	-	-	-	-	45	2.5	.18	296	.11	-	-	1.3
5-27	6.3	-	-	-	-	-	-	63	1	.069	85	.033	-	-	9.5
5-29	7.1	232	8.2	63.5	96	-	-	39	2	.079	104	.021	.12	.079	4.65
6-5	8.1	-	-	-	-	-	-	33	30	.32	338	.13	-	-	2
6-7	6.7	-	-	-	-	-	-	29	50	.35	205	.079	-	-	1.3
6-9	7.1	218	2.5	8.2	99	11.6	66	67.5		.068	59	.023	.28	.09	2.8
6-11	7.1	-	-	-	-	-	-	24	18	.22	83	.032	-	-	3.8
6/12/26	System Failure on Manual Operation														
6-27	6.3	-	-	-	-	7.2	58.4	60.5	7	.34	298	.1	-	.034	6.6
6-30	2.8	-	-	29	99	-	-	73	1	.18	119	.045	.77	-	18
7-3	5.6	289	17.5	29.1	94.5	8.1	86	54	15	.28	210	.08	.27	.07	2.7
7-7	4.9	-	-	13.5	98	2.1	89	28.2	45	.48	277	.11	.19	.015	.86
7-8	5.4	-	-	33.3	96	4.2	90.5	15	40	.36	194	.073	.35	.042	.88
7-9	5.1	-	-	-	-	-	-	50	30	.48	303	.11	-	-	2.7
7-10	5.2	-	-	-	-	-	-	38.2	37	.72	703	.32	-	-	.6
7-11	5.5	-	-	-	-	-	-	24.6	45	.75	425	.16	-	-	.52
7-14 (10PM)	4.8	323	41	49.5	99	5.2	85	54	39	.51	232	.088	.52	.046	3.6
7-15 (6AM)	4.8	357	38	35.5	99	3.33	76	48.6	39	.54	292	.11	.28	.02	1.4
7-15 (2PM)	5.0	340	11.5	10	99	2.5	68	13.5	40	.44	242	.092	.10	.02	.39
7-15 (10PM)	4.9	340	11	5.5	98	4.05	72	37	39	.55	346	.13	.037	.02	.28
7-16 (2PM)	4.1	-	-	1.5	89	8.0	75	-	27	.41	419	.16	.0084	-	-

TABLE 21

Results of Activated Sludge Pilot Plant Operation

Date	Detention Time, Hrs.	Influent NH_3 mg/l	% NH_3 Red.	Infl. Phenol mg/l	% Phenol Red.	Influent CN mg/l	% CN Red.	% COD Red.	Settleable Sol. ml/l	System Solids MLSS Lbs.	 MLVSS Mg/l	 lbs.	Removal LBS/LB MLVSS/Day Phenol	 Cyanide	 COD
7/17/69	12.6	340	10	45	99	2.5	72	-	30	.51	327	.12	.03	.01	-
7/18/69 8am	10.5	-	-	-	-	-	-	-	30	-	-	-	-	-	-
3pm	10.5	-	-	.38	97	1.4	64	-	25	.43	246	.09	.003	.008	-
11pm	10.5	153	11	.38	97	2.0	60	-	23	-	-	-	-	-	-
7/19/69	9.8	-	-	-	-	-	-	-	30	-	-	-	-	-	-
7/20/69	9.6	-	-	-	-	-	-	-	30	-	-	-	-	-	-
7/21/69	9.6	-	-	-	-	-	-	-	24	.33	191	.07	-	-	-
7/22/69	10.1	-	-	-	-	-	-	-	24	.31	220	.08	-	-	-
7/23/69	9.1	-	-	-	-	-	-	-	24	.32	189	.07	-	-	-
7/24/69	9.8	68	0	21.5	99	10.4	93	-	24	.31	140	.05	.39	.17	-
7/25/69	9.8	-	-	-	-	-	-	-	12	.31	250	.09	-	-	-
7/29/69	8.9	-	-	-	-	-	-	-	6	.15	197	.07	-	-	-
7/30/69 7am	8.0	-	-	6.8	97	7.2	94	-	5	.20	289	.11	.07	.07	-
3pm	8.0	-	-	-	-	-	-	-	-	-	-	-	-	-	-
7/31/69 7am	9.9	-	-	-	-	4.0	90	-	9	.18	315	.12	-	.03	-
3pm	9.9	-	-	-	-	2.6	89	-	-	-	-	-	-	-	-
11pm	9.9	-	-	-	-	1.5	76	-	-	-	-	-	-	-	-
8/1/69	15.2	119	14	-	-	2.2	86	-	12	.23	392	.15	-	.008	-
8/2/69	7.2	-	-	-	-	-	-	-	21	-	-	-	-	-	-
8/3/69	17.0	-	-	-	-	-	-	-	28	-	-	-	-	-	-
8/4/69	8.3	271	37	-	-	-	-	-	18	.36	574	.21	-	-	-
8/5/69	6.8	388	16	-	-	5.7	93	36	18	.46	772	.29	-	.02	.82
8/6/69	10.0	350	12	-	-	-	-	29	20	.41	670	.25	-	-	.78
8/7/69	10.0	348	5	-	-	2.0	85	-	20	-	-	-	-	-	-
8/12/69	10.0	-	-	-	-	-	-	14	-	-	-	-	-	-	-
8/13/69 4pm	9.4	-	-	-	-	-	-	14	-	.26	392	.15	-	-	.31
3am	9.3	-	-	-	-	-	-	29	-	.26	410	.15	-	-	.08

Table 22

Results of Activated Sludge Pilot Plant Operation

									System Solids			Removal LBS/LB MLVSS/Day		
Date	Detention Time, Hrs.	Influent NH_3 mg/1	% NH_3 Red.	Infl. Phenol mg/1	% Phenol Red.	Influent CN mg/1	% CN Red.	% COD Red.	MLSS LBS.	MLVSS Mg/1	lbs.	Phenol	Cyanide	COD
8-18 (7A-3P) (Start)	4.15	637	-	3.6	-	13.6	-	-	.53	729	.28	-	-	-
8-18 (3P-11P)	4.15	612	39	36.5	99	8.6	74	55	-	-	.28	.14	.024	1.65
8-18 (11P-7A)	4.15	629	22	34.5	99	6.5	-	70	.50	733	.28	.13	-	2.45
8-19 (7A-3P)	4.15	-	-	-	-	-	-	-	.70	979	.38	-	-	-
8-19 (3P-11P)	4.15	-	-	-	-	-	-	-	.72	1003	.38	-	-	-
8-19 (11P-7A)	4.15	-	-	38	99	2.3	92	46	.77	1084	.41	.09	.0067	.6
8-20 (7A-3P)	3.4	213	-	36	99	19.2	83.5	67	.83	1117	.43	.13	.058	1.9
8-20 (3P-11P)	3.4	-	-	-	-	-	-	-	.70	1120	.43	-	-	1.45
8-20 (11P-7A)	3.4	221	-	30	99	2.3	82	33	.85	1190	.46	.10	.0079	.54
8-21 (7A-3P)	3.4	-	-	-	-	-	-	-	.93	1275	.49	-	-	-
8-21 (3P-11P)	3.4	-	-	-	-	-	-	-	.92	1300	.49	-	-	.79
8-21 (11P-7A)	3.4	187	-	7.3	99	2	85	28	.90	1234	.47	.0024	.0057	.935
												Wasted Sludge		
8-25 (7A-3P)	3.4	626	38	42	97	13.2	98	39	.56	803	.37	.21	.067	1.73
8-25 (3P-11P)	3.4	-	-	-	99	-	-	-	.60	872	.33	-	-	-
8-25 (11P-7A)	3.4	612	2.3	38	98	-	-	49	.64	931	.36	.17	-	1.42
8-27 (7A-3P)	2.4	340	-	2.4	99	6.44	64	17	.88	1210	.46	.16	.027	.52
8-27 (3P-11P)	2.4	-	-	-	-	-	-	-	1.1	1515	.58	-	-	-
8-27 (11P-7A)	2.4	221	4	1.0	92	2.72	44	22	.79	1014	.40	.0075	.0144	.53
9-5 (9:00A-10:00A)	1.94	706	9.8	39	84	3.2	66	41	.6	817	.31	.41	.626	.97
9-5 (10:30A-11:30A)	1.94	-	-	-	-	-	-	-	-	760	.29	-	-	.93
9-5 (12:00-1:00P)	1.94	646	0.0	30	95	1.8	25	11	.56	718	.28	.42	-	.99

Summer temperatures ranged from 18-32°C. These fluctuating temperatures encountered were another variable introduced into the system. Studies by Kostenbader[7] indicated 35°C as the temperature for optimum biological activity.

Sludge Settling

Sludge settling tests to determine clarifier sizing were conducted using the sludge from the trailer mounted pilot activated sludge unit. These tests were carried out in the same column as used in this study of coal washery effluent.

From Table 23 it is evident that adequate suspended solids removal is not complete even after 2 hours of static settling; 450-550 mg/l suspended solids or 12% of the initial solids remain in suspension. This supports the experimental results of the pilot plant studies which showed poor solids retention, sludge washout, and the inability to maintain adequate solids loading. Addition of polyelectrolyte was considered in improving sludge settling. When anionic polymers (Nalco 675 & 673) were added to collection barrels used for storing pilot plant effluent, 70-80% more sludge was collected. No further flocculant tests were conducted. Future studies should be conducted in this area if the optimization of temperature, pH and feed concentration does not result in good sludge settling.

Oxygen Demand

As indicated in the Waste and Sources Section, the combined coke plant effluent may contain reducing agents exerting an immediate oxygen demand (IDOD) on the biological system. Several aeration tests were run in order to check the IDOD of the waste and to determine oxygen transfer rates. Tests were conducted in a 55 gallon drum. A constant air pressure was supplied through a 1/2 inch pipe. Since the oxygen transfer to the waste was compared to that of plant water, aeration rate, air pressure, and liquid depth were kept constant. No attempts were made to measure the absolute values of air flow, oxygen uptake rate, or aeration efficiency.

Aeration tests were conducted by filling the drum to a constant level (6 inches from the top) and checking dissolved oxygen content. The DO of the mill supply water was close to zero when collected. The addition of small quantities of sodium sulfite reduced the DO to zero at the start of aeration.

A galvanic cell DO meter (Yellow Springs Instrument), calibrated on city water by the Winkler method of DO determination, was used for monitoring DO levels in the barrel. Aeration was continued at constant air pressure until DO concentration approached saturation.

TABLE 23

Activated Sludge Settling Tests

Test No. 1. MLSS = 4435 mg/l

Mg/l Suspended Solids/Percent Removal Versus

Sampling Point	Column Depth (ft)	Settling Time, Minutes				
		10	20	45	60	120
1	.5	1214/72	921/79	780/82	634/86	493/89
2	2.7	1827/59	964/78	562/87	626/86	534/88
3	4.7	3674/17	952/79	645/86	652/86	551/88
4	6.7	3422/12	2165/51	684/85	562/87	522/88

Test No. 2. MLSS = 3022 mg/l

Mg/l Suspended Solids/Percent Removal Versus

Sampling Point	Column Depth (ft)	Settling Time, Minutes				
		10	20	45	60	120
1	.5	1203/60	886/71	602/80	591/80	450/85
2	2.7	1280/58	961/68	744/75	641/79	540/82
3	4.7	1778/41	793/70	608/80	592/80	549/82
4	6.7	2486/18	839/72	632/78	666/78	563/81

Time of aeration, water and air temperature, and barometric pressure were also measured. Oxygen saturation values for water at prevailing temperatures and pressures were obtained from handbook tables. Values of DO in the barrel, in terms of mg/l below saturation, were plotted on semilog paper against aeration time (Figure 8).

The figure indicates that on occasion the waste taken at Sampling Point C-7A exerts IDOD. On August 7, one test showed no evidence of IDOD as indicated by an immediate increase in DO concentration at the beginning of aeration. In the second test, however, it was necessary to aerate for 35 minutes before a constant increase in DO was observed. Similar results were observed on August 8 when an aeration period of 2 hours was required before a constant increase was recorded.

The curves for oxygen transfer, as shown on Figure 8, may be used for calculation of the relative oxygen transfer coefficient, alpha. Alpha is defined as the ratio of oxygen transfer rates to waste and to water (K_{La} waste $\div$ K_{La} water) using standard methods of calculation as described in literature.[1] It was found that K_{La} for the raw mill water supply on August 6 was equal to 7.67 $Hr.^{-1}$ at 28°C, and that the K_{La} for the waste from C-7A, on August 8, was 6.65 $Hr.^{-1}$ at the same temperature. The corresponding alpha value is 0.86. This was the highest value observed, as can be seen from Figure 8 where the lines for the rest of the tests all have slopes of greater negative values. The samples from C-7A which exhibited an IDOD also showed lower Alpha values once a constant rise in DO was reached.

Complete Mix Activated Sludge Unit

The next step planned for the biological testing was to test the applicability of a complete mix activated sludge system to the treatment of the total Coke Plant effluent flow at sample point C-7A. This series was started on October 16, 1969. One of the six aeration compartments of the pilot unit was isolated and the piping was arranged so that it received both the influent waste stream and the recirculated sludge from the secondary settling tank. The testing employed a five hour aeration time and the same supply system as used in the previous activated sludge tests. Since this test started in mid-October and the operating temperatures were extremely low, removal efficiencies for phenol and cyanide were low as compared to the previous plug flow system results. It was felt that most of the problem could be traced to the low temperature. After approximately two weeks of unsatisfactory operation, the pilot plant testing was discontinued and future testing was performed with small scale laboratory equipment.

During the week of November 10-14, a small-scale laboratory activated sludge unit was set up as a complete mix system. The

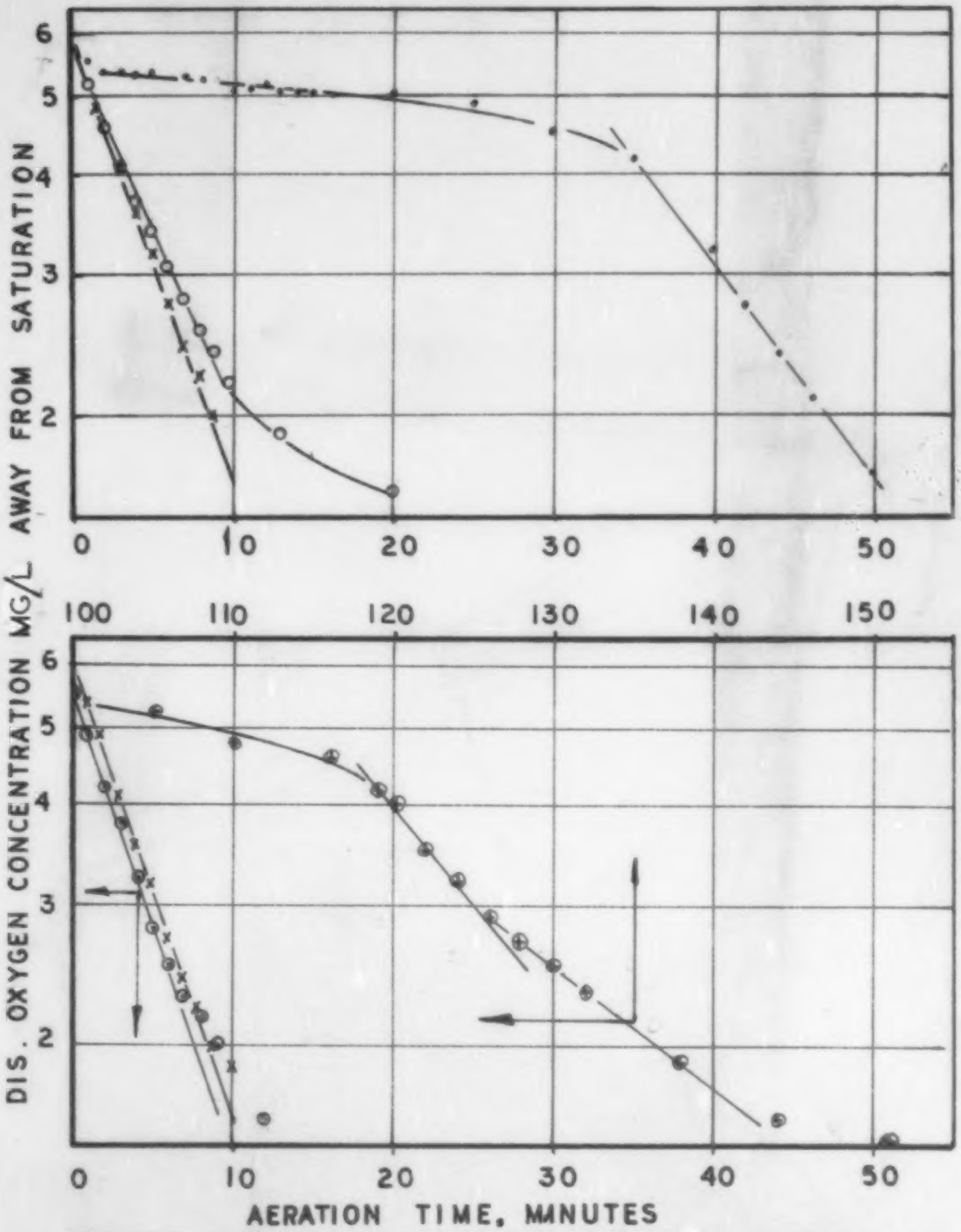


FIG. 8
AERATION OF MILL WATER AND COKE PLANT EFFL.

TEST	DATE	SOURCE	TEMP. DEG. C.
—×—	8-6	MILL SUPPLY	28
—·—	8-7	C-7 A	37
—o—	8-7	C-7 A	33
—●—	8-8	C-7 A	37
—o—	8-8	C-7 A	37

activated sludge unit was set up as a complete mix system. The unit, which was developed by Bio Development Associates of Austin, Texas, is shown in Figure 9. It was constructed out of plexiglass and contained a settling compartment with an effluent weir which was separated from the aeration chamber by an adjustable baffle. An air diffuser stone was positioned in the aeration chamber such that the rolling action of the mixed liquor drew settled solids from the settling compartment under the adjustable baffle and back into the aeration chamber. There was provision for wasting solids from the settling compartment. A Masterfex tubing pump was used for supply and provided excellent control over the feed rate. The air supply was adjusted to give a minimum of 4.0 mg/l DO in the mixed liquor as well as a good rolling action. The unit had a total volume of 10 liters, two liters of which composed the settling compartment.

In addition to eliminating the variable introduced by temperature fluctuation it was decided to supplement the influent feed of total coke plant effluent with phenol to circumvent the problem of low phenol concentration encountered in the pilot plant tests. The nitrogen content of the waste was sufficient to promote biological growth. Phosphoric acid provided both nutrient phophorus and pH buffering. The pH of this waste was adjusted in this manner in order to run for at least seven days to avoid shocking the microorganisms by frequent changes in the waste material. The laboratory unit was seeded with acclimated sludge from the pilot plant.

Detention times were varied from 3-8 hours by adjusting the pump feed rate. Biological solids increased slowly, but significantly, during the initial period of operations at a detention time of 8 hours. When flow was increased to reduce the detention time to 3-4 hours, phenol removal efficiency was quite poor (58-62%). Data are presented in Table 24. Following this series of tests the sludge from the unit was frozen and used for another series shown on Table 25. Results of these two series showed that good phenol removal is possible at high concentrations with detention times of 8 hours. Erratic removal efficiency was evident at lower detention times indicating no positive trend. The results indicate that further study of the complete mix system as well as the plug flow system in a larger scale pilot plant should be considered.

Sludge Growth and Oxygen Uptake

Several attempts were made to determine oxygen requirements and rates of sludge growth using laboratory equipment. Sludge growth, in terms of pounds per pound of phenol feed, was very high as shown in Table 25. It is realized that it is not possible to have an increase in material by simple conversion of organic substances, such as phenol to biological sludge. It is possible of course

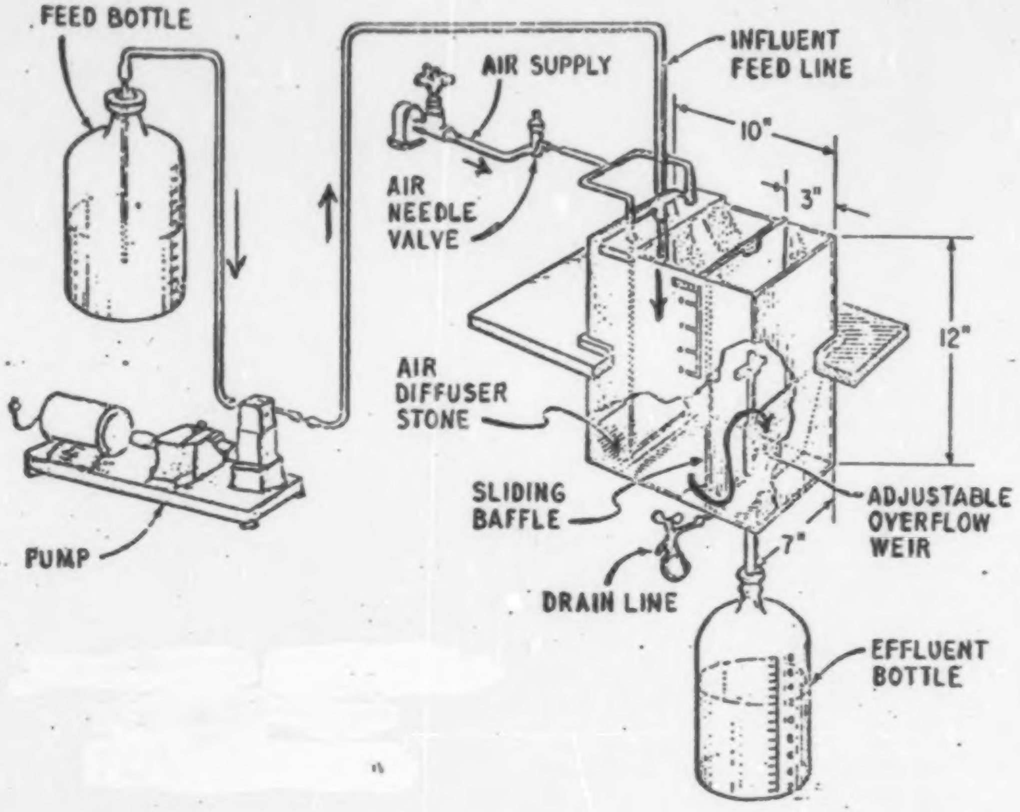

FEED BOTTLE
AIR SUPPLY
INFLUENT FEED LINE
10"
3"
AIR NEEDLE VALVE
12"
AIR DIFFUSER STONE
SLIDING BAFFLE
ADJUSTABLE OVERFLOW WEIR
7"
PUMP
DRAIN LINE
EFFLUENT BOTTLE

Figure 9. LABORATORY SCALE ACTIVATED SLUDGE UNIT FLOW DIAGRAM

TABLE 24

Laboratory Biological Studies

				Phenol			Cyanide		
				Mg/1			Mg/1		
Date	pH In	Detention Time Hours	MLVSS	Influent	Effluent	% Reduction	Influent	Effluent	% Reduction
11/10/69									
8:00 AM	7.3	8	-	67.5	40.8	40	-	-	-
3:00 PM	8.0	8	-	64	29.7	54	-	-	-
11/11/69									
5:00 PM	8.0	8	-	96.5	.9	99	-	-	-
12:00 PM	8.0	8	603	75.5	1.0	98	-	-	-
11/12/69									
9:00 AM	7.3	8	721	78	.53	99	7.52	2.56	64
5:00 PM	-	8	677	61.3	2.0	97	-	-	-
12:00 AM	-	8	-	71.6	22.0	70	-	-	-
11/13/69									
9:00 AM	8.0	3	-	62.6	23.7	62	13.3	6.4	52
4:00 AM	8.0	3	-	62.6	26.2	58	9.7	5.9	40
11/14/69									
4:30 AM	8.0	3	1,192 wasted sludge	54.6	1.34	97	-	-	-
11/17/69									
8:00 AM	8.0	4.5	857	-	-	-	-	-	-
4:00 PM	8.0	4.5	1,081	-	-	-	-	-	-
12:00 AM	8.0	4.5	1,152	105.6	5.32	95	-	-	-
11/18/69									
4:00 PM	8.0	3	1,250	96.2	49.4	49	31.4	11.4	64
12:00 AM	8.0	3	1,180	83.6	45	46	-	-	-
12/2/69									
2-8 PM	-	6.3	-	148	52.7	57.5	-	-	-
3-8 AM	-	6.3	788	141	34	76	-	6.2	-
3-9 PM	-	8.0	-	102.5	.3	99	-	-	-
4-1:30 PM	-	8.0	688	98	7.1	93	-	-	-
5-5:30 AM		8.0	1200	99	.2	99	-	-	-
5-2:30 PM		6.0	1200	81.5	10.7	87	-	-	-
5-9 PM		6.0	-	70.4	12.2	81.5	-	-	-
6-2:30 AM		6.0	-	65.6	15.5	76	-	-	-
6-8:30 AM		6.0	1200	-	11.4	-	-	-	-
7-8:30 AM		6.0	1492	wasted2.6 liters mixed liquor			-	-	-
7-8:30 PM		6.0	-	25.8	-	-	-	-	-
8-8:30 AM		6.0	940	81	27	60	-	-	-
8-3:00 PM		6.0	948	77.6	35.2	55	-	-	-
10-7 PM		7 hr	-	94.6	41.6	56	-	-	-
11-7 AM		7 hr	1720	92.3	57.2	38	-	-	-
12-7 AM		7 hr	-	85.8	6.2	94	12	4.5	-
12-3 PM		6.3 hr	-	116.6	.9	99	-	-	-
12-8:30 PM		6.3 hr	-	193.3	.4	99	-	-	-
15-8:30 AM		6.3	600	151	.8	-	-	-	-
16-8:30 AM		4.0	372	-	.3	-	-	-	-
16-6:30 PM		4.0	-	187.78	43.2	77	-	-	-
16-9:15 PM		4.0	-	195.9	66.7	64	-	-	-
17-3 AM		4.0	-	213.6	90.5	53	-	-	-
16-8:30 AM		8.0	800	193.8	.3	99	-	-	-
16-6:30 PM		8.0	-	447	.3	99	-	-	-
16-11 PM		8.0	-	476	.5	99	-	-	-
17-5 AM		8.0	-	445.8	.5	99	-	-	-
17-12:30 PM		8.0	224	456.2	.9	99	-	-	-
17-3 PM		6.0	209	248	209	47.5	-	-	-
17-8 PM		6.0	-	140.8	35.2	81.5	-	-	-
17-9 PM		8.0	-	397.6	.4	99	-	-	-
18-12 AM		8.0	-	208.4	.4	99	-	-	-
18-3 AM		6.0	-	194.8	42.4	79	-	-	-
18-3 AM		8.0	360	413.6	.5	99	10.1	2.8	-
18-6 AM		6.0	540	206.4	.5	90.5	10.1	1.8	-

TABLE 25

Results of Lab Tests Sludge Growth and Oxygen Uptake

Date	Time	Mixed Liquor Volume	Phenol Added Grams*	MLVSS			Lbs Lb Phenol	Lbs Lb COD	Oxygen mg/1 Min	Uptake mg Hr-g VSS
				mg/1	Grams	Increase Grams*				
9-18	2:30 P	4200	0	718	3.0	0	-	-	.28	22.7
	8:00 P	4958	.193	710	3.5	0.5	2.6	.33	-	-
9-19	12:00 A	5550	.154	701	3.9	.38	2.5	.38	-	-
	7:00 A	6600	.268	743	4.9	1.0	3.7	.52	-	-
	9:30 A	6900	0	785	5.4	0.5	-	-	.75	57.5
	10:20 A	3600	0	1300	4.7	-	-	-	1.21	46

REMARKS: * Since last time of monitoring
** Volume reduced by decanting
Feed Solution 256 mg/1 phenol 1672 mg/1 COD

that other materials, beside phenols, enter into the bio-synthesis process. Sludge growth in terms of pounds per pound of COD, therefore, may be more meaningful.

Since complete removal of suspended solids from the mixed liquor had proved to be incomplete at 2 hrs. settling time, it was decided to investigate sludge growth rate in a batch system. Aeration was carried out in a 3 gallon glass bottle, calibrated in liters. The tests were run with acclimatized sludge obtained from a feed reservoir prepared by fortifying a sample from sampling point C-7A with analytical grade phenol to a final concentration of 256 mg/l. The volume of mixed liquor was increased over a 20 hour period by sample additions. Samples were withdrawn and analyzed for phenol, cyanide, soluble COD, and VSS. COD concentration in the feed solution was 1672 mg/l.

It is evident that the bacteria developed in the activated sludge plant will readily consume analytical grade phenol when oxygen and nutrients are in ample supply. Sludge buildup rates under these conditions may be as high as 3.7 pounds of volatile suspended solids per pound of phenol consumed. In terms of pounds per pound of COD, however, the growth rate may go as high as .52.

Another attempt to measure sludge growth was made in a continuous flow, completely-mixed laboratory unit operated at high phenol loading rates. Results of these tests are shown in Table 26.

Taking the increase in MLVSS during the test period (12-8 to 12-11) the average sludge growth over a 72 hour period amounted to 14.6 grams in the 10 liter volume of the system. The average phenol feed during that same period was 8 grams. This is equivalent to a sludge growth of 1.8 lbs of sludge per lb of phenol, which is about half of the rate found in the batch system. The reason for this lower value could be loss of sludge in the overflow. Complete separation of activated sludge was not accomplished during the first two days of the test.

Oxygen uptake was calculated from plots of DO reduction with time. Respiration rates were obtained by temporarily increasing the dissolved oxygen content of the mixed liquor to approximately 5 mg/l and by monitoring the rate of DO reduction in the mixed liquor after turning off the oxgyen supply. Data obtained from this study is plotted on Figure 10.

Assuming that the combined flow to the planned activated sludge treatment process is 2000 gpm and that 5 hours retention time is required, the oxygen required based on the above respiration rates will amount to approximately 311 lbs/hr at an MLVSS of

TABLE 26

Results of Compl Mixed Lab Unit, Activated Sludge System
6.35 Hr Detention Time

Date	Time	Phenol mg/l In	Phenol mg/l Out	% Reduction	MLVSS mg/l	Cyanide mg/l In	Cyanide mg/l Out
12-8	2:00 PM	81.5	27	67	948	-	-
	5:00 PM	77.6	35.2	55	940	-	-
12-9	7:00 AM	88.2	48.2	45	1180	43.9	20.1
	7:00 PM	89.7	47.9	46.5	-	20.4	8.6
12-10	8:00 AM	86.2	37.2	59	908	-	-
	2:00 PM	94.6	41.6	58.5	1132	-	-
	7:00 PM	-	-	-	-	-	-
12-11	7:00 AM	85.8	6.2	93	2406	-	-
	6:00 PM	116.6	.91	99	-	-	-

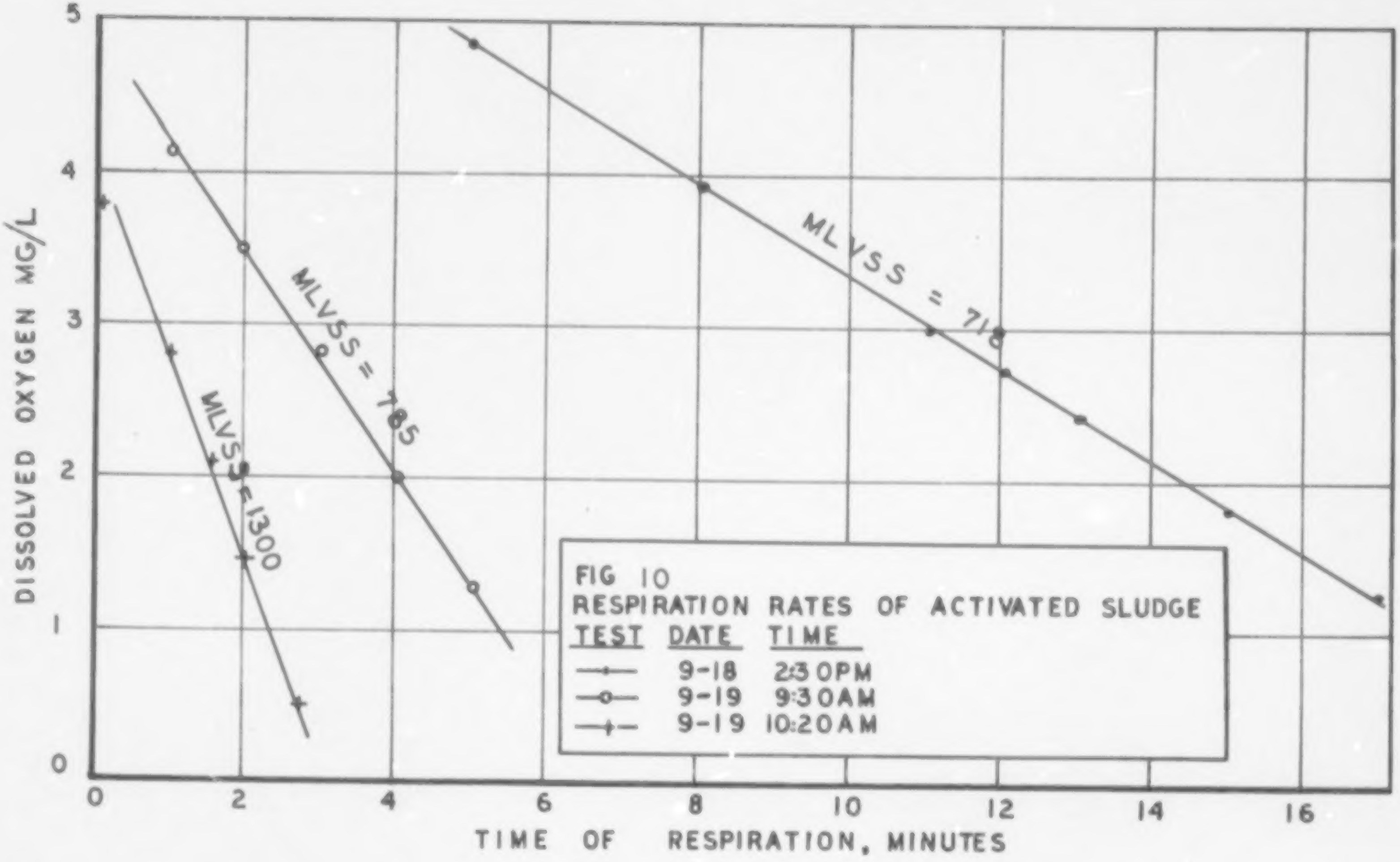


FIG 10
RESPIRATION RATES OF ACTIVATED SLUDGE

1000 mg/1.

Data From Table 25

Average Oxygen Uptake/Minute = 0.8 mg/1

Average Volume of Mixed Liquor = 4.9 Liters

Flow = 2000 GPM

MLVSS = 1000 mg/1

Detention Time Needed = 5 Hours

$$\text{Lbs. Oxygen/Hour} = 0.8 \times \frac{2000 \times 4.9}{1000 \times 454} \times 300 \times 60$$

$$\text{Lbs. Oxygen/Hour} = 0.8 \times \frac{9800}{454{,}000} \times 300 \times 60$$

$$\text{Lbs. Oxygen/Hour} = 0.8 \times .0216 \times 300 \times 60 = \underline{\underline{311}}$$

A series of tests designed to produce additional operating data was run from January 5 through February 7 in two parallel units with different feed rates. Phenol concentrations were not supplemented in this series. These results are shown in Tables 27 and 28.

On January 15, sludge from CF&I's sanitary sewage plant was added to one of the units; the other unit was taken out of service. A detention time of 5 hours was maintained during the remainder of the test. Excellent sludge growth was ob-

TABLE 27

Laboratory Biological Studies

Date	pH In	Detention Time Hrs.	Mixed Liquor VSS (mg/l)	Phenol mg/l Influent	Phenol mg/l Effluent	% Reduced	Cyanide mg/l Influent	Cyanide mg/l Effluent	% Reduced
1/6/70									
10:30 am#1	8.0	11	-	51.7	1.4	97			
10:30 am#2	8.0	10	-	51.7	27.8	46			
8:00 pm#1	8.0	11	-	279	46	83			
8:00 pm#2	8.0	10	-	279	48	83			
1/7/70									
4:00 pm#1	8.0	9	-	138	.26	99			
4:00 pm#2	8.0	11	-	138	.17	99			
12:00 am#1	8.0	8.8	-	148	3.6	98			
12:00 am#2	8.0	11	-	148	.16	99			
1/8/70									
8:00 pm#1	8.0	6.3	-	8.2	.33	96			
2:30 am#2	7.2	8.2	-	8.4	.53	94			
1/12/70									
8:30 am#1	8.0	6.6	-	56	.06	99			
8:30 am#2	8.0	22.0	-	56	.01	99			
5:00 pm#1	8.0	6.6	-	-	.12	-	3	.5	84
5:00 pm#2	8.0	8.2	-	-	.1	-	3	.5	84
6:00 am#1	8.0	6.3	-	-	.10	-	-	.4	-
6:00 am#2	8.0	8.2	-	-	.03	-	-	.5	-
1/13/70									
9:30 am#1	8.0	5.1	212	32.8	10.3	70	36	13.8	62
9:30 am#2	8.0	8.2	220	32.8	.33	99	36	9.3	74
1/14/70									
8:00 am#1	8.0	5.1	176	32.8	10.7	68	36	12.6	65
8:00 am#2	8.0	8.2	260	32.8	.14	99	36	10.8	70
11:00 pm#1	8.0	5.1	-	37.9	.96	98	30.5	16.6	45
11:00 pm#2	8.0	8.2	-	37.9	.4	99	30.5	8.8	71
1/15/70									
8:00 am#1	8.0	5.1	-	37.9	.37	99	30.5	13	57
8:00 am#2	8.0	8.2	-	37.9	.39	99	30.5	12.8	58
8:00 pm#1	8.0	5.1	-	37.9	.3	99	30.5	15.5	49
8:00 pm#2	8.0	8.2	-	37.9	.36	99	30.5	15.0	51
1/16/70									
10:00 am	8.0	5.1	-	-	-	-	-	-	-
1/17/70									
10:00 am	8.0	5.1	-	-	-	-	-	-	-
1/18/70									
10:00 am	8.0	5.1	-	-	-	-	-	-	-
1/19/70									
8:00 am	8.0	5.1	-	37.9	.11	99	-	-	-
1/20/70									
8:00 am	8.0	5.1	-	-	-	-	-	-	-
5:00 pm	8.0	5.1	1,044	28.3	.18	99	-	-	-

TABLE 28

Laboratory Biological Studies

Date	pH In	Detention Time Hrs.	Mixed Liquor VSS (mg/l)	Phenol mg/l Influent	Phenol mg/l Effluent	% Reduced	Cyanide mg/l Influent	Cyanide mg/l Effluent	% Reduced
1/21/70									
7:30 am	8.0	5.1	1,215	-	-	-	-	-	-
4:00 pm	8.0	5.1	-	50.8	.01	99	28.8	23.2	19
11:30 pm	8.0	5.1	-	-	.14	-	-	22.7	-
1/22/70									
7:30 am	8.0	5.1	1,391	28.9	.10	99	-	-	-
10:30 am	8.0	5.1	-	29.4	.12	99	20.6	19.2	7
3:30 pm	8.0	5.1		28.1	.01	99	21.5	10.9	49
1/23/70									
7:30 am	8.0	5.1	1,787	-	-	-	-	-	-
8:00 pm	8.0	5.1	-	-	-	-	-	4.4	-
1:30 am	8.0	5.1	-	-	.13	-	-	-	-
1/24/70									
9:00 am	8.0	5.1	1,666	-	-	-	-	-	-
1/25/70									
9:00 am	8.0	5.1	1,137	-	-	-	-	-	-
1/26/70									
7:30 am	8.0	6.5	1,732	29	.13	99	8.7	2.9	67
1/27/70									
7:30 am	8.0	5.1	1,964	-	.1	99	22.2	7.0	69
1/28/70									
7:30 am	8.0	5.1	1,824	32.7	.10	99	23.2	9.8	58
1/29/70									
7:30 am	8.0	5.1	2,239	-	.13	-	-	-	-
1/30/70									
7:30 am	8.0	5.1	2,265	-	.14	-	-	-	-
1/31/70									
9:00 am	8.0	5.1	3,149 wasted sludge	-	-	-	-	-	-
2/1/70									
9:00 am	8.0	5.1	1,440 wasted sludge	-	-	-	-	-	-
2/2/70									
2:30 am	8.0	5.1	1,523	-	-	-	-	-	-
12:00 am	8.0	5.1	wasted sludge	36.8	.19	99	21	8.4	60
2/3/70									
9:00 am	8.0	5.1	2,599	-	-	-	-	-	-
2/4/70									
11:00 pm	8.0	5.1	-	15.9	.1	99	35	6.2	82
2/6/70									
11:00 pm	8.0	5.1	-	207	.19	99	35	5.0	86
2/7/70		SLUDGE TRANSFERRED TO TRICKLING FILTER PILOT UNIT							

tained and the sludge exhibited good settling tendencies with very little carry-over. Phenol removals were generally over 99%.

Since cyanide removals generally were only 50-55%, the bench scale unit previously removed from service was reactivated and placed in series with the operating unit to see if additional cyanide removal could be obtained. The sludge characteristics changed completely from that of the unit receiving the raw feed. The color of the sludge went from brown to tan and the floc formed was very light and exhibited very poor settling. It became necessary to collect the total effluent, and decant to retrieve the sludge which was returned to the unit. Cyanide removal could not be realistically determined because of the variations in the MLSS.

A general evaluation of performance of the biological systems studied is difficult because of wide fluctuations in feed concentrations of phenol and cyanides resulting in great variation in biological activity. In particular, sludge growth rate was practically never constant. It is believed that this latter problem was largely due to the poor settling characteristics of the sludge and the consequent loss of solids from the system.

Attempts were made to correlate phenol removal with detention time and effluent phenol concentrations. Fiqures 11 and 12 present this correlation. Figure 11 shows the effect of aeration tank detention time on phenol removal. The data points shown were taken from the operation of the pilot unit when MLVSS varied between 800 and 1300 mg/l and when the system appeared to operate without any interference. Data is shown in Tables 27 and 28.

Although there is considerable scattering of data, it appears that the pilot plant unit (plug flow system) required a minimum of 3 hours detention time. Ninety nine percent removal may be expected with detention times above 4 hours.

Phenol removal (Lbs/Lb MLVSS/Day was calculated and plotted against corresponding influent phenol concentrations (Figure 12). The values used were taken from pilot plant operations when phenol feed was above 10 mg/l and when the system seemed to be free of interferences. Data is shown in Tables 27 and 28. Because of the scattering of data and because of the limited number of points available, no attempt was made to insert a straight line, which is expected from such plots. The slope of such a line would be equal to K (slope of lbs Ø-OH removed vs time).[14]

The test data obtained from the pilot and laboratory studies indicates that it is feasible to treat CF&I coke plant wastes with an

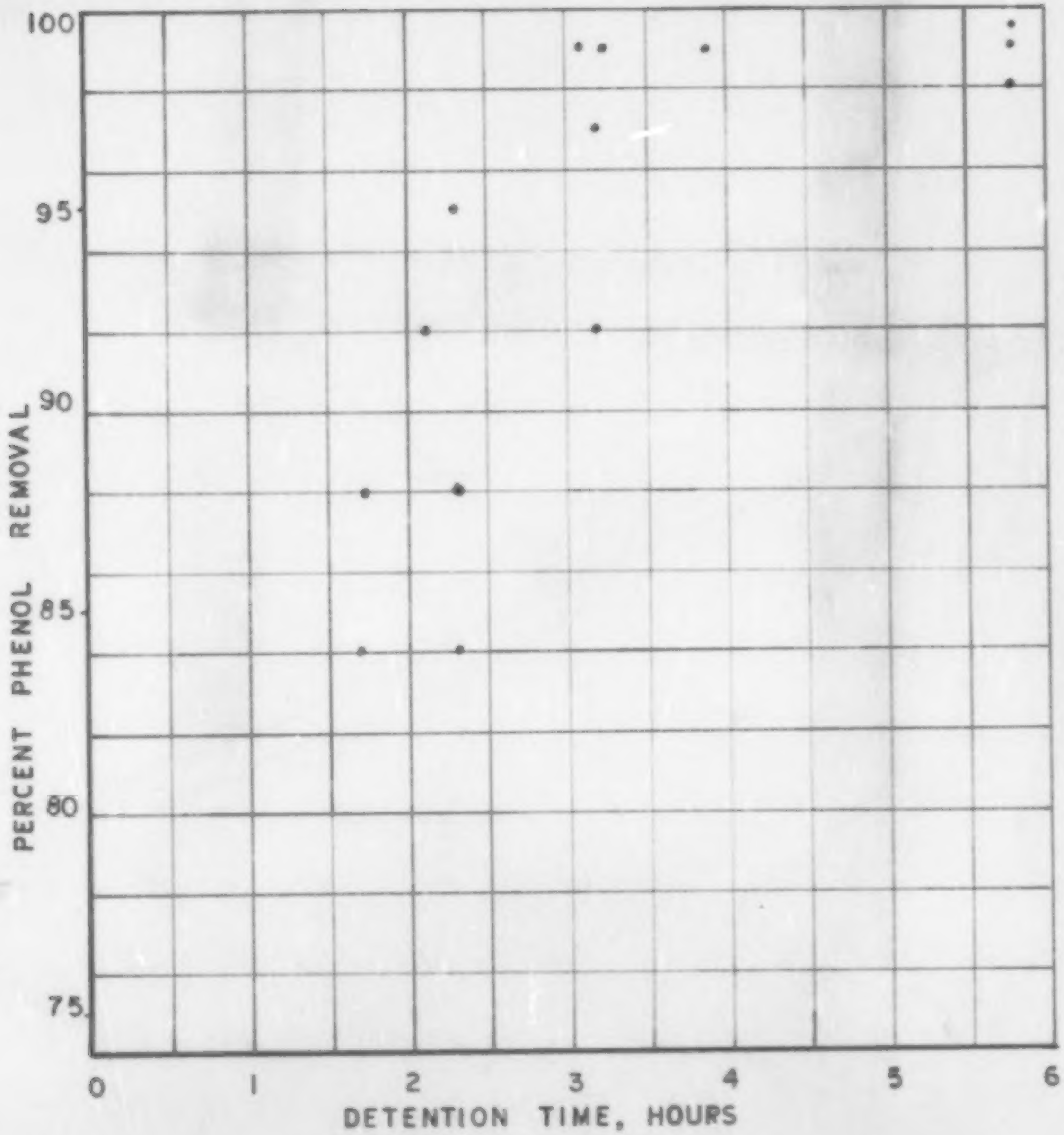


FIG. 11
PHENOL REMOVAL VS. DETENTION TIME

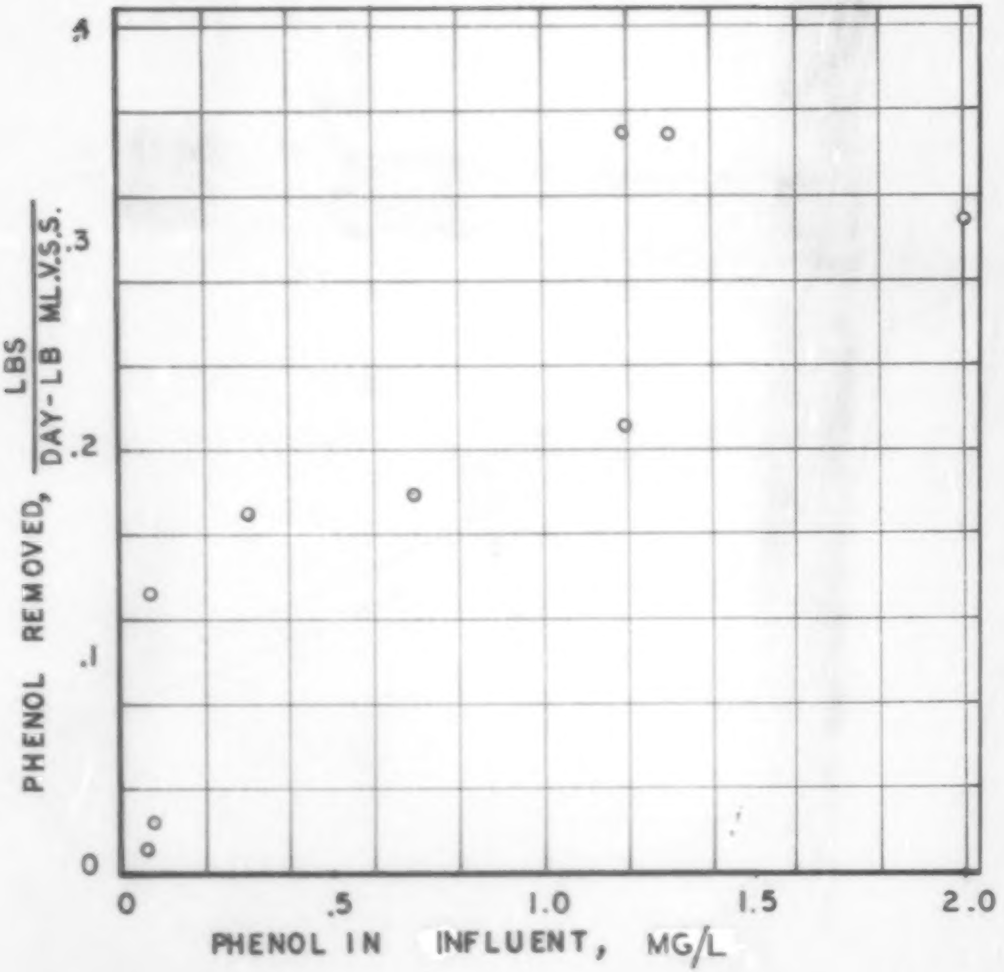


FIG 12
PHENOL REMOVAL VS. EFFLUENT CONCENTRATION

activated sludge system. For removal of phenol, cyanide and other organic materials, no significant ammonia or color reduction can be expected. However, before an actual biological activated sludge system could be installed, the problems encountered during the study must be resolved and additional data obtained to identify the operating parameters of such a system. This can be accomplished through the installation and operation of a larger pilot unit over a 6 months period. Nalco Chemical Company has recommended that a pilot activated sludge unit, with a capacity of treating 20,000 gallons per day, be installed for this purpose. They further recommend that a package sewage plant such as that available from Chicago Pump Division of FMC Corporation, Chicago, Illinois, be utilized for this purpose. The cost of such a system is approximately $15,000. Rental of such equipment is also possible.

The pilot system would include a primary settling basin large enough to provide a 24 hour detention time and provisions for heating the aeration tank. A basin of this size is needed to reduce the effects of wide fluctuations in temperature and pollutant concentrations which were partially responsible for the erratic data obtained in the present study. An additional benefit would be the reduction of suspended solids levels, especially in the case of flotable tars which Kostenbader and Flecksteiner[7] found to be troublesome in a biological treatment system. Oil skimming facilities would be a part of this installation.

The holding basin would allow the pilot plant to operate on a continuous basis even though no influent samples were taken from the waste stream during the coal washery operation. The other piping and operative changes recommended in this report to reduce the total flow and load to the final treatment plant should be completed before this testing.

Increasing the phenol concentration would enhance the possibilities for more consistent and efficient operation of an activated sludge biological system. Removal of uncontaminated streams from the combined effluent would be a step in this direction. On the other hand, if the phenol-contaminated streams contain materials which are toxic to the microorganisms, these materials would also be present in larger concentrations, along with the bio-degradable phenolic materials, in the reduced volume effluent. Test results would soon show any toxic effects and dilution water could be added if necessary. Work done by Ludberg and Nicks[15]; Hepner, Tallon and Fisher[16]; and Kostenbader and Flecksteiner[7] showed that dilutions of 1:1 are necessary in treating various coke plant wastes biologically. Most of these tests were made on the more concentrated flushing liquor wastes.

Several parameters to be used for the operation of the pilot system can be defined. The optimum temperatures for successful biological treatment of coke plant wastes as reported by Abson and Todhunter[17] are 25-30°C. Kostenbader and Flecksteiner[7] suggest that 21-38°C is preferred. Some heat may have to be added to the coke plant waste to maintain the waste stream temperature in these ranges. Optimum pH ranges as reported by Abson and Todhunter[17] for the mixed liquor are 7.0 - 8.0 for phenol oxidation and 6.5 - 7.0 for cyanide oxidation. Good phenol removals were obtained at mixed liquor pH of 6.9 - 7.7. This indicates that a pH of 7.0 would be the most desirable to maintain. Spent sulfuric or hydrochloric pickle liquor could be used for pH adjustment.

Test data from the pilot and laboratory studies showed that a detention time of 3.4 hours in the aeration basin would provide 99% phenol removal and 70-80% reduction in cyanides. Some operating activated sludge treatment systems are using from 9-19 hours (Kostenbader and Flecksteiner[7], Ludberg and Nicks[15]). Detention times in the planned pilot system will be varied to determine the optimum time necessary for good phenol and cyanide removal.

The operation of this pilot unit will provide the data necessary for the design of a treatment plant to handle the combined coke plant effluent.

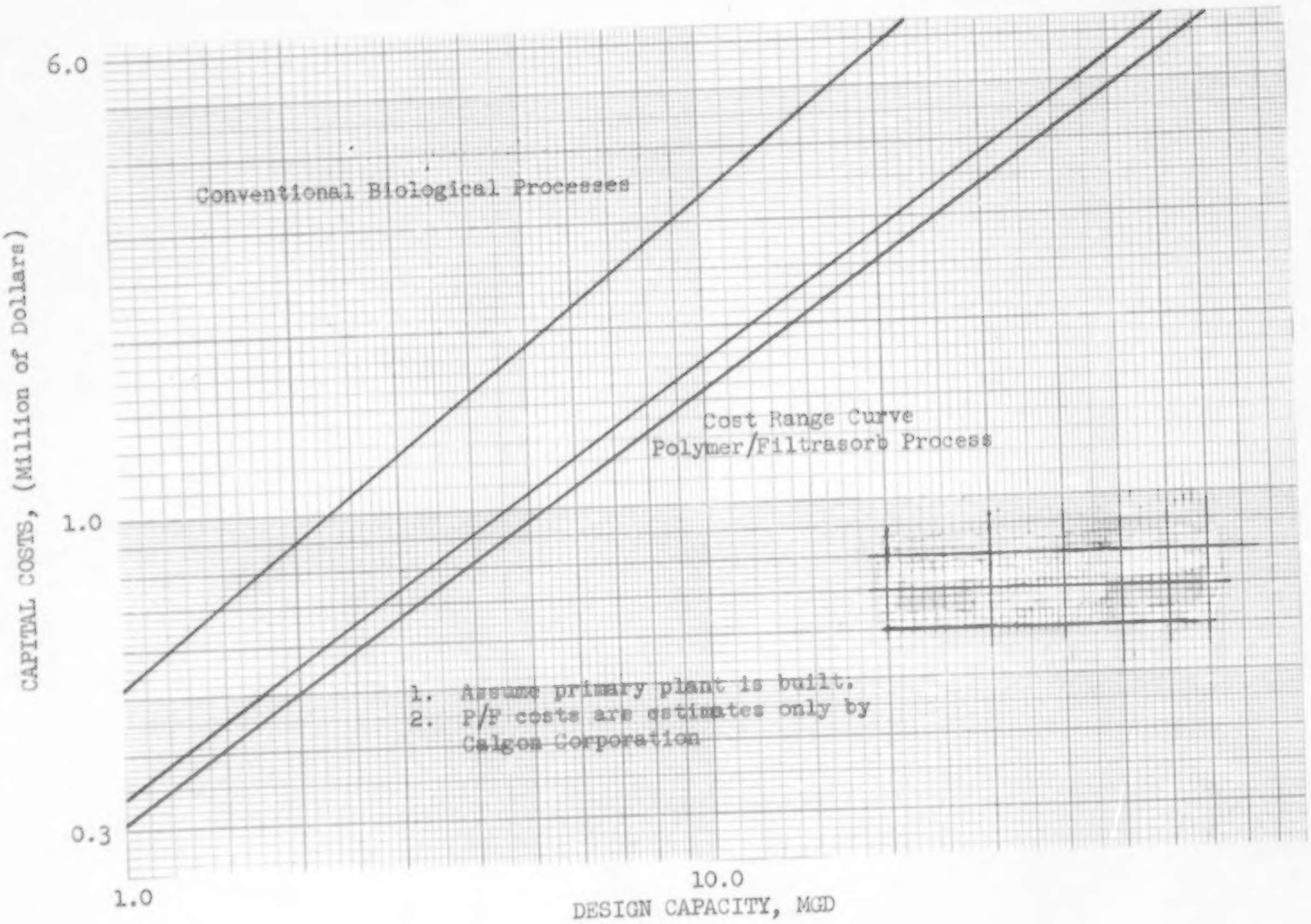
Figure 13 shows the capital cost of a conventional biological municipal treatment plant at various flows. At a flow of 2,000 gpm or 2.9 MGD, capital costs for the coke plant wastes would approximate $1,050,000.

Trickling Filter

Another series of tests involving biological oxidation of the coke plant ammonia flushing liquor was conducted using a trickling filter pilot unit. The tests were originally planned at the site of the combined coke plant waste stream at sampling point C-7A, but low temperatures necessitated moving the unit inside. The distance from C-7A made it impractical to use that waste source for the study. Ammonia flushing liquor diluted with raw mill water was substituted since it was available near the pilot installation.

The trickling filter system is shown in Figure 14. Its components are:

1) A 12 stack (Dow Surfpack) tower, 20 inches x 20 inches x 23 feet high constructed of 14 gauge steel plate. A removable pan perforated at the bottom was used to distribute the feed to the media.



1. Assume primary plant is built.
2. P/F costs are estimates only by Calgon Corporation

Figure 13. CAPITAL COST COMPARISON

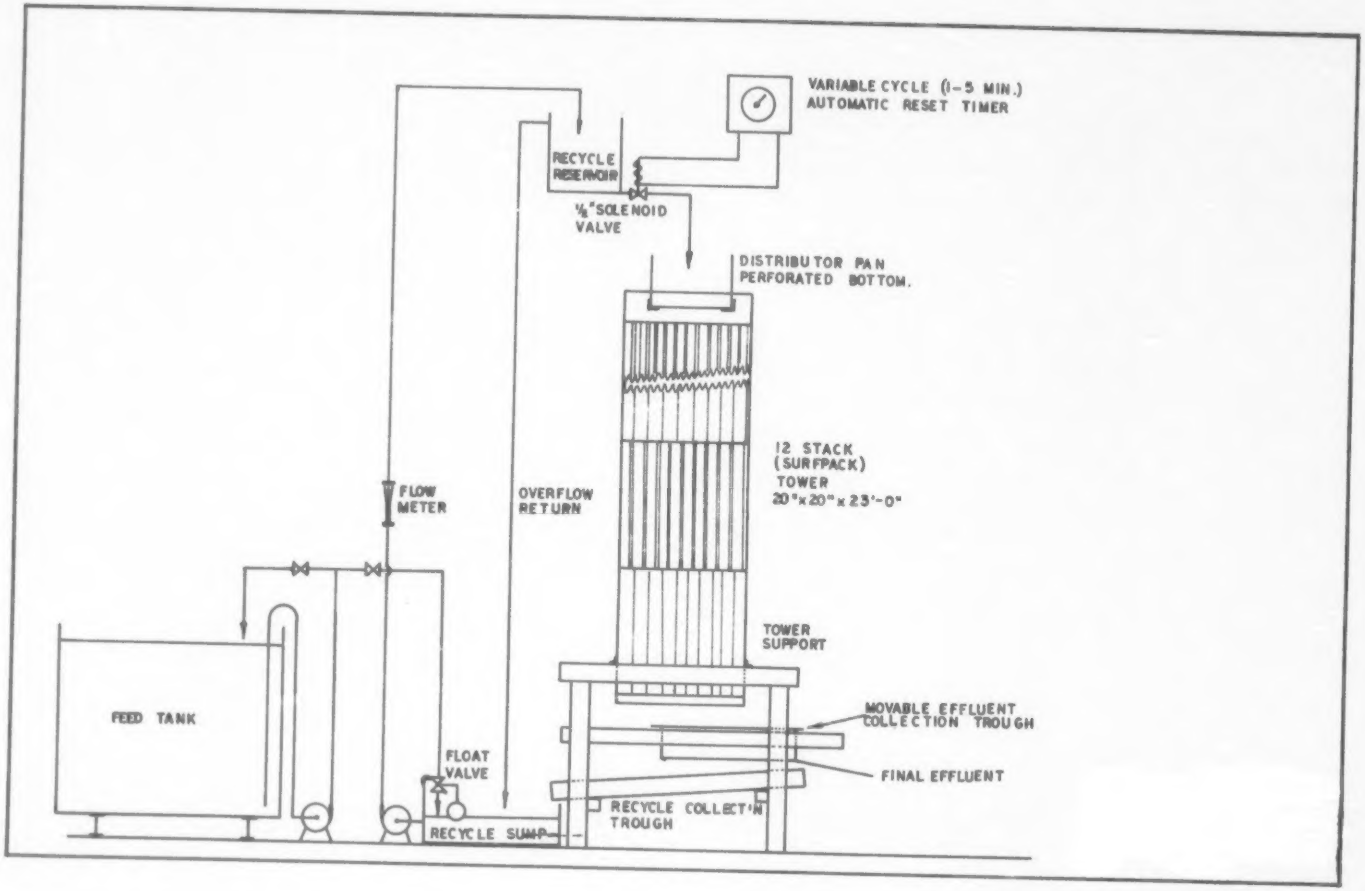

VARIABLE CYCLE (1-5 MIN.)
AUTOMATIC RESET TIMER
RECYCLE RESERVOIR
½" SOLENOID VALVE
DISTRIBUTOR PAN PERFORATED BOTTOM.
12 STACK (SURFPACK) TOWER 20"x20"x23'-0"
FLOW METER
OVERFLOW RETURN
TOWER SUPPORT
MOVABLE EFFLUENT COLLECTION TROUGH
FINAL EFFLUENT
FEED TANK
FLOAT VALVE
RECYCLE SUMP
RECYCLE COLLECT'N TROUGH

Figure 14. TRICKLING FILTER PILOT PLANT LAY-OUT

2) 67.5 cubic feet of plastic trickling filter media. (Dow Surfpack)

3) Two holding tanks, 943 gallon capacity. One tank was used to store and cool the ammonia flushing liquor. The other served as a primary feed tank where the ammonia flushing liquor was diluted with fresh mill water, pH adjustments were made, and phosphate nutrient was added.

4) One half of a 55 gallon drum cut vertically was used as a recycle sump. The level in the sump was maintained by a float-operated valve connected to the primary feed tank system.

5) A 55 gallon drum cut in half horizontally was used as a feed reservoir at the top of the filter. Flow from the reservoir through a solenoid valve was controlled by an automatic, variable cycle, reset timer. Excess feed overflowed back to the recycle sump.

6) Three pumps were required for the system. One pump, 20-30 gpm capacity at 20 ft head, transferred ammonia flushing liquor to the primary feed tank on a batch basis. Another pump continuously recycled the primary feed to provide make-up feed for the recycle sump. This pump was designed for 5-10 gpm at 30 ft head. The third pump, designed for 5-10 gpm flow at 30 ft head, transferred the feed from the recycle sump to the feed reservoir at the top of the filter.

The activated sludge used in the laboratory complete-mix study was used to start the filter. Sludge from CF&I's sanitary sewage plant system was also added. The unit was initially fed a mixture of 8 parts fresh mill supply water to 1 part ammonia flushing liquor.

One hundred percent recycling at 1.0 gpm was practiced at the start in order to develop a biologically-active slime. Approximately 1 week elapsed before any activity was noticed. At this time the recycle was reduced to 90% with a 10% flow through the system. Various rates of recycle were used to obtain data on phenol and cyanide removals. Tables 29-30 show the results of the trickling filter test.

From the data obtained, it is apparent that a trickling filter would not remove the phenol as readily nor as completely as the activated sludge system even at high dilution of the flushing liquor. Average removals obtained were 69%. In addition only 33% reduction of cyanides was obtained as compared to 75-80% in the activated sludge system.

TABLE 29

Results of Trickling Filter Tests

	pH		Phenol Mg/l		%	Cyanide Mg/l		%	Flow		Dilution
	In	Out	In	Out	Reduced	In	Out	Reduced	Recycle GPM	To Sewer GPM	H$_2$O FL
2/9/70											
Start 9am											
6:00 pm (Grab)		8.0	392	364	-	20.5	9.0	-	1.0	0	8 - 1
11:15 pm "		"	404	364	-	18.5	5.5	-	-	0	
2/10/70											
3:45 pm "		8.0	390	390	-	-	6.5	-	1.0	0	8 - 1
11:15 pm "		"	-	394	-	-	5.5		-	0	
2/11/70											
4:00 pm "		8.0	411	399	-	14.5	6.5	-	1.0	0	8 - 1
11:15 pm "		"	417	402	-	14	6.0		-	0	
2/12/70											
11:15 pm "		8.0	400	374		4.0	4.5		1.0	0	8 - 1
2/13/70											
3:00 pm "		8.0	-	255		-	3.5		1.0	0	8 - 1
2/16/70											
10:00 am "		8.0	241	212		-	-		1.0	0	8 - 1
2/17/70											
8:30 am "		8.0	242	176		-	-		1.0	0	8 - 1
7:15 pm "			245	182	-	-	-	-			
2/18/70											
7:30 am "		8.0	250	6.2	98	-	-	-	.09	0.1	8 - 1
2/19/70											
7:30 am "		8.0	-	-	-	-	-	-	.09	0.1	8 - 1
2/20/70											
7:30 am "		8.0	-	-	-	-	-	-	1	-	8 - 1
2/21/70											
7:30 am "		8.0	-	-	-	-	-	-	1	-	8 - 1
2/22/70											
7:30 am "		8.0	-	-	-	-	-	-	1	-	8 - 1
2/23/70											
7:30 am "		8.0	-	17	-	-	-		.09	0.1	8 - 1

TABLE 30

Results of Trickling Filter Tests

	pH		Phenol Mg/1		%	Cyanide Mg/1		%	Flow		Dilution	
	In	Out	In	Out	Reduced	In	Out	Reduced	Recycle GPM	To Sewer GPM	H_2O	FL
2/24/70 7:30 am	-	8.0	-	7.5	-	-	-	-	.85	.15	8 -	1
2/25/70 7:30 am	-	8.0	-	-	-	-	-		1.35	.25	8 -	1
2/26/70 8:30 am	-	8.0	-	26.2	-	-	-		1.45	.15	8 -	1
2/27/70 8:15 am	-	8.0	-	1.1	-	-	3.3		1.35	.25	8 -	1
2/28/70 8:00 am	-	8.0	203.5	27.5	87	-	-	-	1.35	.25	8 -	1
3/1/70 8:00 am	-	8.0	187.5	29.2	85	-	-	-	1.35	.25	8 -	1
3/2/70 7:30 am	7.8	7.9	-	2.4	-	4.2	2.8	33	1.2	.40	4 -	1
3/3/70 8:00 am	8.1	8.1	231	114	51	14.3	12.3	14	1.2	.40	4 -	1
3/4/70 8:00 am	7.65	8.05	190	14	93	9.0	8.3	8	1.2	.40	4 -	1
3/5/70 8:00 am	8.25	8.0	175	55	68	11.5	6.3	45	1.2	.40	4 -	1
3/6/70 8:00 am	8.4	8.4	-		-	-	-	-	1.2	.40	4 -	1
12:00 pm	8.4	-	-		-	-	-	-	1.27	.33	4 -	1
3/7/70 8:00 am	8.0	8.4	171.5	66.5	61	7.8	5.5	30	1.27	.33	4 -	1
3/8/70 8:00 am	8.0	8.5	222	136	39	16.0	5.6	65	1.27	.33	4 -	1
3/9/70 8:00 am	8.0	8.0	185	115	38			-	1.27	.33	4 -	1

ADSORPTION

Activated Carbon

A portion of the research studies was designed to determine the adsorption capabilities of activated carbon on coke plant waste water, using Calgon Corporation's Filtrasorb 300 and 400 granular activated carbon. Physical properties and specifications for the two types of carbon are included in Appendix G, page 266. Three separate test runs were conducted during the study. For the first two tests, waste water was pumped from sampling point C-7A. Samples were not taken during the operation of the coal washery to avoid undue solids loadings. Test 3 was conducted on ammonia flushing liquor from sample point C-13. These two streams were chosen in order to determine the adsorption wave front and carbon organic capacity as a function of both the highest and the lowest dissolved organic levels which require treatment (See Tables 6-10 for waste characteristics). Interpolation of the results can provide information relative to intermediate dissolved organic levels. During the testing there was no attempt to manipulate either the temperature or pH of the waste water. In addition to testing for the adsorption of the organics in an 18 ft x 5 inch diameter bed of granular activated carbon, the second and third runs also tested a new process recently developed by Calgon Corporation utilizing granular activated carbon, a catalyst, and oxygen for destruction of cyanide.

The 18 ft x 5 inch activated carbon adsorption bed was divided into a 3 ft bed followed by three 5 ft beds in series. Figure 15 is a photograph of this equipment and Figure 16 is a schematic layout. The cyanide destruction equipment is pictured in Figure 17 and shown schematically in Figure 18. The two cyanide reaction columns were each 1 ft in diameter by 6 ft high (internal dimension). Each column was packed with a 4 ft bed of activated carbon (Filtrasorb 400). The cyanide destruction unit was operated on the effluent from the adsorption columns.

Equipment procured for the test included:

1) A 1½ HP Self-Priming pump, capacity 30-50 gpm at a 60 ft head, to transfer the waste water from the sewer to a holding or storage tank.

2) A 6 ft diameter x 2 ft galvanized tank, capacity 423 gallons, for storage of untreated waste water.

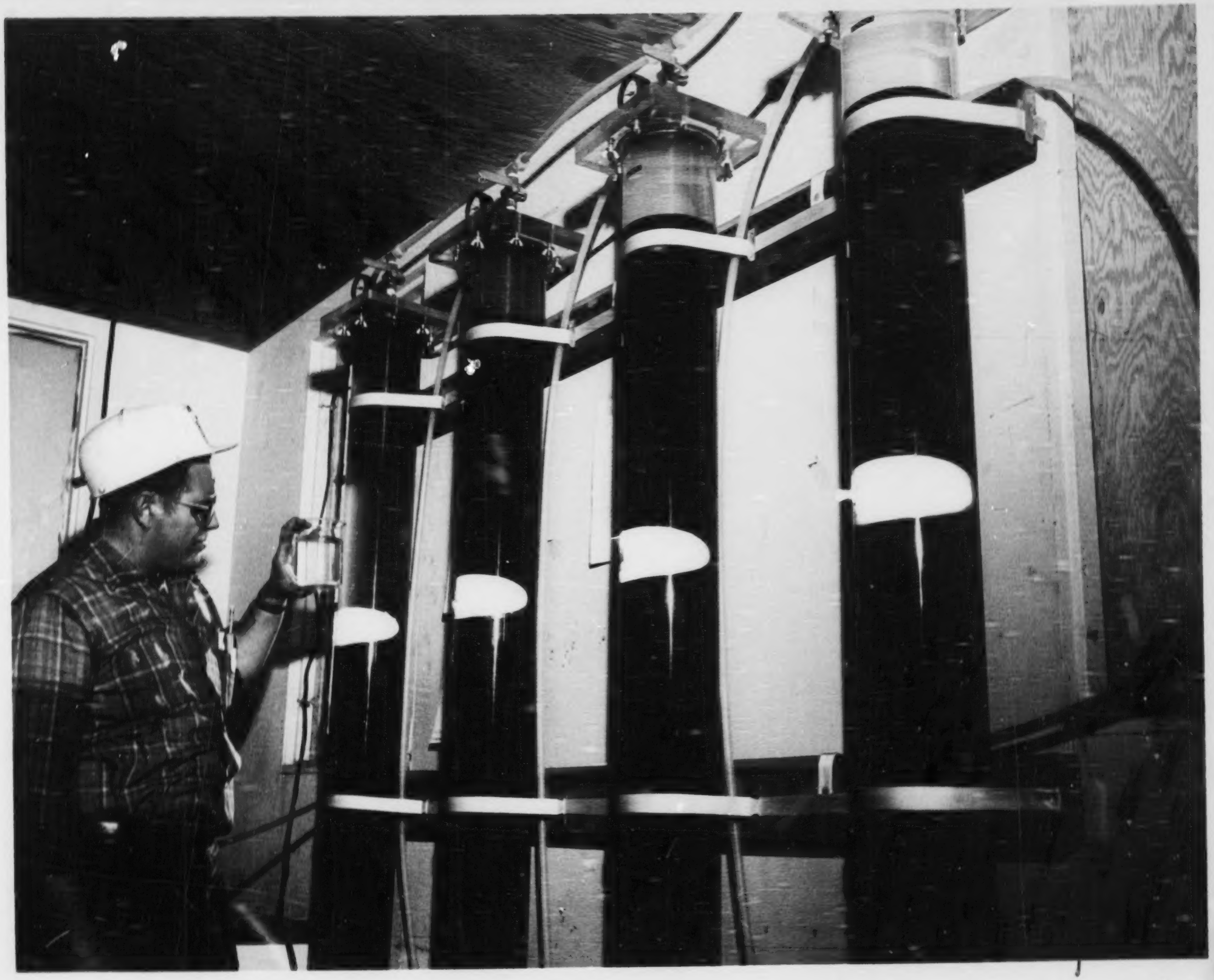


Figure 15. PHOTOGRAPH OF PILOT CARBON ADSORPTION COLUMNS

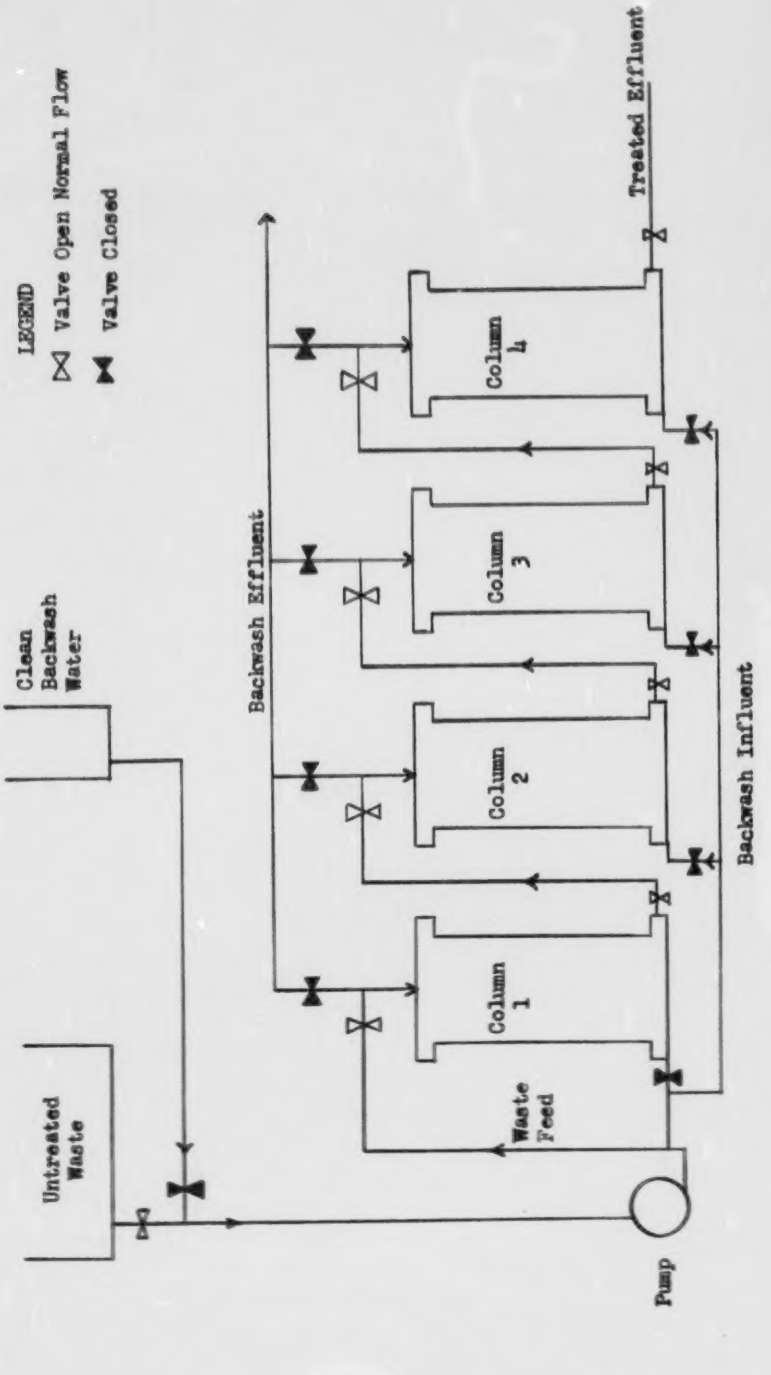

LEGEND
⧖ Valve Open Normal Flow
⧗ Valve Closed
Clean Backwash Water
Untreated Waste
Pump
Waste Feed
Column 1
Column 2
Column 3
Column 4
Backwash Effluent
Backwash Influent
Treated Effluent

Figure 16. PILOT CARBON COLUMNS

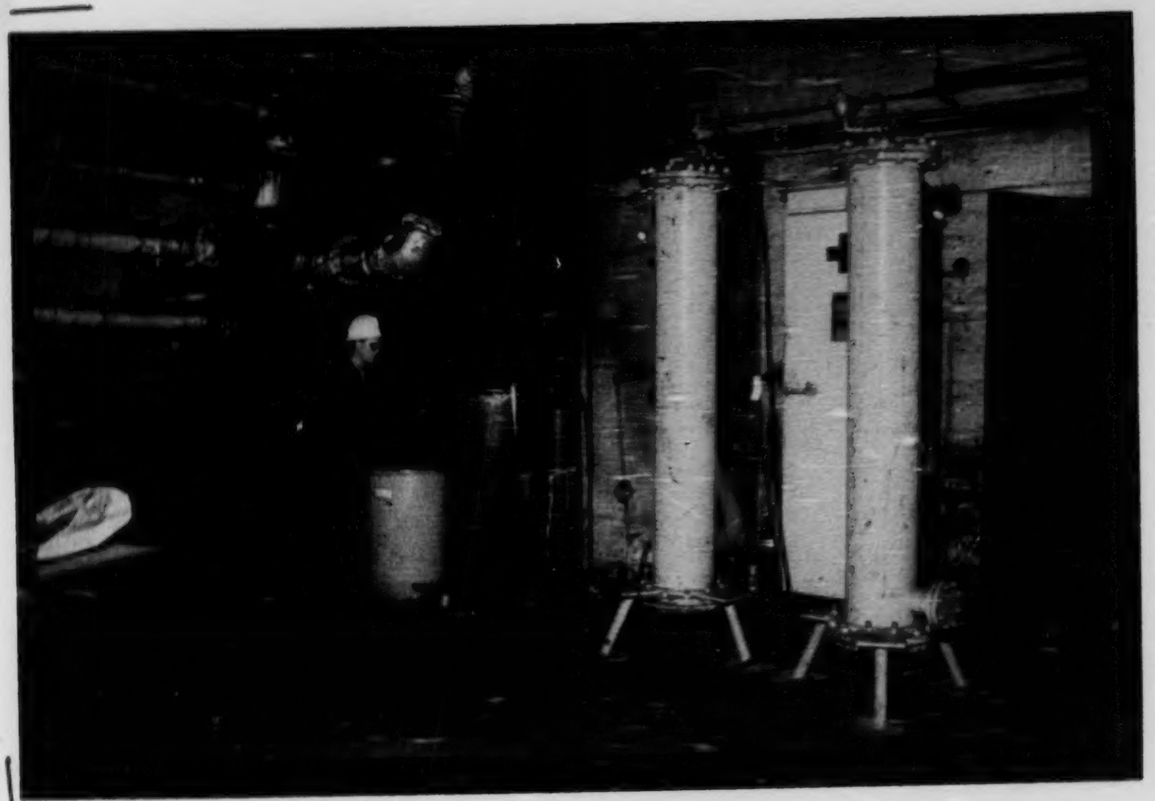


Figure 17. PHOTOGRAPH OF CN DESTRUCTION EQUIPMENT

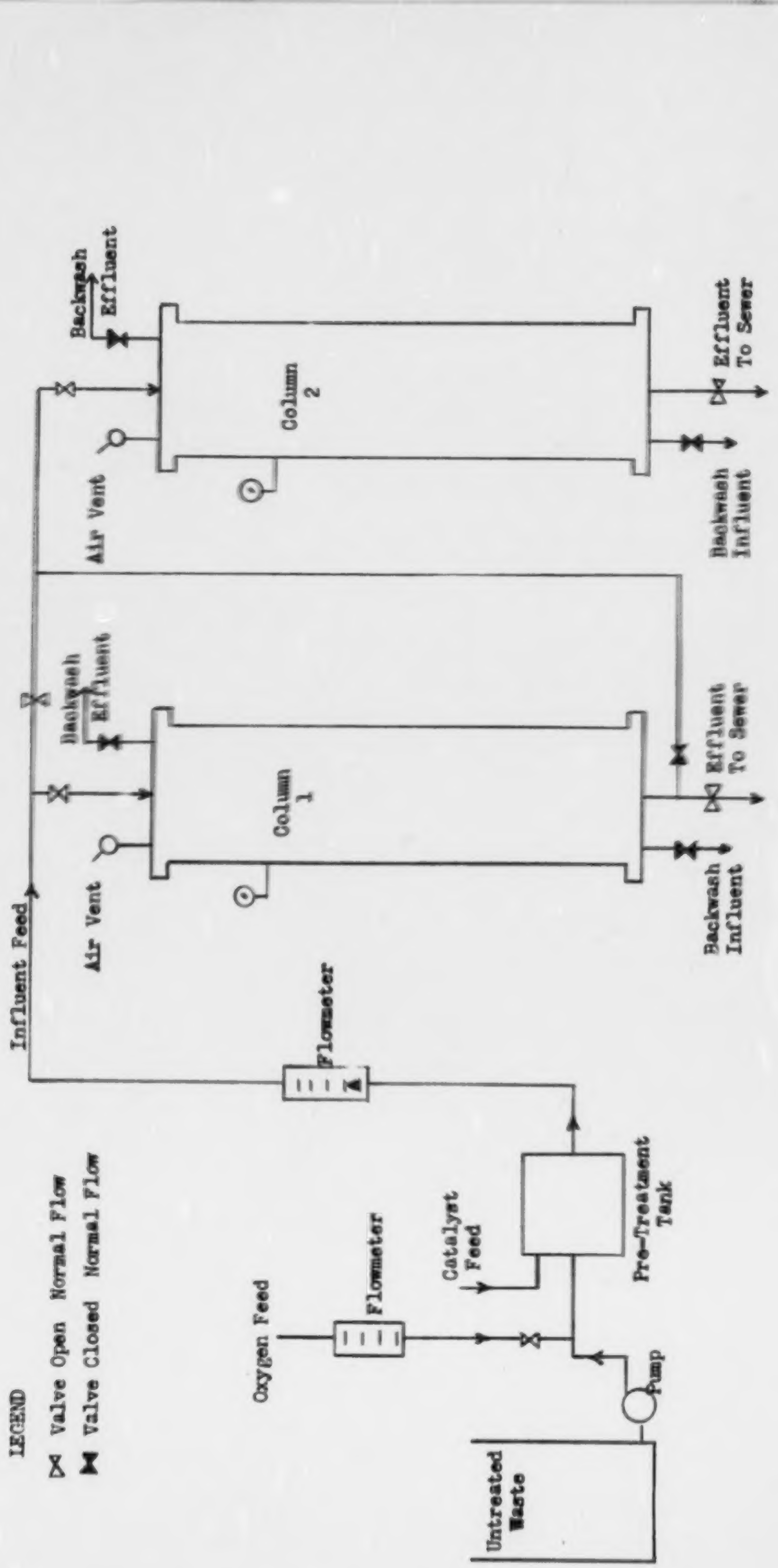

LEGEND
⧖ Valve Open Normal Flow
⧓ Valve Closed Normal Flow
Untreated Waste
Pump
Oxygen Feed
Flowmeter
Catalyst Feed
Pre-Treatment Tank
Flowmeter
Influent Feed
Air Vent
Column 1
Backwash Effluent
Backwash Influent
Effluent To Sewer
Air Vent
Column 2
Backwash Effluent
Backwash Influent
Effluent To Sewer

Figure 18. CYANIDE DESTRUCTION COLUMNS

3) A 9 ft diameter x 2 ft galvanized tank, capacity 952 gallons, for storage of water treated by the carbon columns. This provided a source of water for the cyanide destruction units.

4) Four transparent, plastic columns 6 ft high x 5 inch diameter (capacity 6.18 gallons) for phenol, color, and organic removal. Removal plates at the top and bottom of the columns provided access for carbon addition and contain ports for inlet and outlet piping, backwashing, and gauge connections.

5) Two stainless steel columns 8 ft 2 inches high x 1 ft in diameter, capacity 44.1 gallons for cyanide removal. Removal plates provide access for carbon additions and contain ports for inlet and outlet piping, backwashing and vent connections. Gauges were mounted on the side of the column.

6) Two 1/2 HP centrifugal pumps, capacity 5-10 gpm at 40 ft head, to provide steady flow and constant water pressure to the treatment columns. These pumps were also used as backwash pumps.

7) A 20 gallon treatment tank for the cyanide removal system. This was a pressure type tank with the necessary fittings for water-oxygen mixture inflow, catalyst injection and mixture outflow to the cyanide removal columns. Oxygen was injected into the untreated water just ahead of this tank.

8) One 45 gallon-capacity polyethylene tank for catalyst preparation, complete with an adjustable chemical feed pump.

9) Oxygen pressure regulator and flowmeter capable of flow adjustments from 0-3.5 SCFH. Commercially pure bottled oxygen was used for oxygen supply.

10) Necessary hoses, pipes and fittings for interconnection of equipment.

Flow Measurement

Water flow through the equipment was measured directly at effluent points by timing the flow into a graduated breaker. Flow was controlled by manipulation of a gate or ball valve downstream from the columns. Flows were monitored three times each 8 hour shift on Monday through Friday and at least twice on Saturday and Sunday. Actual flows for each test series are tabulated on Tables 31 and 34. By utilizing a daily backwash procedure, relatively stable pressure drop and flow were maintained. Catalyst flow was controlled by the adjustment of the metering pump and checked by catalyst tank

TEST 1 ACTIVATED CARBON

SAMPLE POINT C-7A

Date 1969	Hours Operated	Flow Rate Gal/Min	Gal/Day	Cumulative Gallons Total
10/27	15.0	.25	225	225
10/28	22.5	.25	337	562
10/29	22.5	.25	337	899
10/30	23.3	.25	350	1249
10/31	23.6	.25	356	1605
11/1	23.5	.25	353	1958
11/2	23.7	.25	356	2314
11/3	23.5	.25	353	2667
11/4	23.5	.25	353	3020
11/5	23.5	.25	353	3373
11/6	22.7	.25	341	3714
11/7	23.5	.25	353	4067
11/8	23.2	.25	348	4415
11/9	23.6	.25	354	4769
11/10	23.5	.25	353	5122
11/11	23.5	.25	353	5475
11/12	23.5	.25	353	5828
11/13	23.5	.25	353	6181
11/14	11.5	.25	173	6354
11/15	23.3	.25	350	6704
11/16	23.6	.25	354	7058
11/17	23.5	.25	353	7411
11/18	0	0	0	7411
11/19	0	0	0	7411
11/20	13.0	.25	195	7606
11/21	23.0	.25	345	7951
11/22	23.5	.25	353	8304
11/23	21.5	.25	323	8627
11/24	23.5	.25	353	8980
11/25	15.0	.25	225	9205
11/26	20.3	.25	305	9510
11/27	23.5	.25	353	9863
11/28	12.0	.25	180	10043
11/29	13.5	.25	203	10246
11/30	24.0	.25	360	10606
12/1	20.5	.25	308	10914
12/2	22.0	.25	330	11244
12/3	22.0	.25	330	11574
12/4	22.2	.25	333	11907
12/5	18.3	.25	275	12182
12/6	22.5	.25	338	12520
12/7	22.0	.25	330	12850
12/8	23.3	.25	350	13200
12/9	20.3	.25	305	13505
12/10	15.7	.25	236	13741

Table 31 Activated Carbon Pilot Operational Data

measurements. Oxygen flow was measured by a calibrated flow meter.

Sampling Program

During the first test grab samples were taken and either analyzed individually or were composited. Individual samples were drawn for phenol, COD, TOC. Samples taken for color, pH and TSS were composited. The samples were fixed, as necessary, according to FWPCA guidelines. Sampling points used provided data to and from each of the four operating columns to determine when each column reached the breakthrough point.

Test two employed composite sampling throughout and included cyanide analyses as well as those tests previously mentioned. Daily samples were composited by mixing 8-hour grab samples.

For test three, grab samples were taken once a day and analyzed for phenol, COD, TOC, pH, color and cyanide.

Test 1

This test was designed to provide operating and design parameters for a treatment plant using activated carbon to remove color, phenol and other organics. A fixed-bed-in-series adsorber configuration (Figure 16), was used containing 60 pounds of activated carbon (Calgon Corporation-Filtrasorb 300). The carbon was divided into 4 columns, a 3 ft bed followed by three 5 ft beds and was operated downflow at 1/4 gpm. Waste water was pumped into the columns at 7-9 psi and 1/4 gpm. The project was scheduled on a 24 hour, 7 day a week basis interrupted only by daily backwashing. Backwashing became necessary to remove suspended solids which caused high pressure differentials and loss of flow through the columns. The test run proceeded until all four columns exhibited breakthrough of phenol.

The results obtained for test one are compiled in Tables 32-33. It can be seen that excellent phenol removal was obtained with an effluent concentration of about 0.01 mg/l. Good reductions of color, COD, and TOC were also obtained. In interpreting this data it must be remembered that the test was designed primarily to obtain design data by operating the columns to the breakthrough point, and consequently the removal efficiency decreased as the test progressed. TOC data was used to determine the adsorption wave front necessary to achieve the levels of phenol, COD and TOC as shown in the initial stages of Test 1 (Table 32). The wave front is the number of feet of bed between the point of saturation of the carbon and the point at which the lowest concentration or breakthrough of the contaminant can be detected. The TOC parameter

TABLE 32

Test 1

Carbon Columns Data

Date	Color % Trans.					Phenol (mg/l)					COD (mg/l)					TOC (mg/l)				
	TK	1	2	3	4	TK	1	2	3	4	TK	1	2	3	4	TK	1	2	3	4
10/27/69																				
2:30 pm	55	64	74	80	84	12	.01	.01	.01	.01	212	36	36	-	-	67	15	7	5	10
10:00 pm	55	67	72	78	79	9.8	.01	.01	.01	.01	280	80	16	-	-	67	17	15	27	12
9:00 am	16	78	90	89	89	12.3	.01	.01	.01	.01	392	-	-	8	16	85	18	10	9	10
2:30 pm	53	84	99	98	99	10.8	.01	.01	.01	.01	198	48	24	16	-	86	17	36	29	18
10:30 pm	48	30	99	100	100	12.4	.15	.01	.01	.01	186	-	28	-	-	81	24	33	25	18
6:30 am	28	35	85	99	98	10.5	.11	.01	.01	.01	298	-	16	16	-	57	24	19	17	12
10/29/69																				
11:30 am	38	68	86	91	95	13.1	.10	.01	.01	.01	368	72	32	32	8	90	27	23	27	32
10:30 pm	42	53	75	78	83	12.2	.12	.01	.01	.01	284	80	72	72	52	92	35	32	30	32
6:30 am	36	38	69	76	83	15.9	.15	.01	.01	.01	48	200	48	64	60	88	40	32	35	33
10/30/69																				
2:30 pm	5	51	91	94	93	29.0	.10	.01	.01	.01	348	123	101	89	131	130	43	33	41	36
10:30 pm	33	56	86	93	95	27.0	.12	.01	.01	.01	392	118	83	91	95	122	41	40	29	31
6:30 am	27	13	64	73	78	27.2	.34	.01	.01	.01	-	-	-	-	-	108	57	36	44	42
10/31/69																				
10:30 pm	22	14	68	76	80	20.6	.23	.01	.01	.01	211	100	-	62	-	119	40	46	49	50
6:30 am	30	11	59	73	70	21.5	.30	.01	.01	.01	-	-	-	-	-	140	52	45	47	52
11/1/69																				
1:30 am	53	70	79	82	82	31.5	.73	.01	.01	.01	244	129	119	112	103	112	24	21	22	23
11/3/69																				
7:00 am	61	75	83	87	87	36.3	5.2	.01	.01	.01	-	-	-	-	-	-	-	-	-	-
6:30 pm	19	47	86	76	92	31.5	12.6	.01	.01	.01	424	65	53	41	29	137	58	25	22	21
11/4/69																				
6:30 am	15	35	65	72	77	15.6	15.3	.08	.01	.01	278	206	59	118	112	160	145	28	23	24
11/5/69																				
6:30 am	24	31	69	76	75	34.8	25.3	.08	.01	.01	378	252	28	53	28	148	129	38	32	31
11/6/69																				
6:30 am	53	-	77	85	90	17.0	-	.08	.01	.01	207	-	61	28	12	91	-	42	22	21
11/7/69																				
6:30 am	40	-	81	91	95	18.0	-	.05	.01	.01	220	-	49	45	53	71	-	26	24	28
11/8/69																				
7:00 am	65	-	87	83	94	24.0	-	4.0	.90	.01	-	-	-	-	-	80	-	29	26	23
11/9/69																				
8:45 am	77	-	97	98	100	25.0	-	4.6	.01	.01	212	-	69	90	20	64	-	22	24	22
11/10/69																				
6:00 am	49	-	78	81	85	24.0	-	.75	.01	.01	270	-	87	71	71	78	-	35	29	29
11/11/69																				
6:30 am	53	-	79	80	85	21.4	-	.20	.01	.01	395	-	75	85	95	111	-	38	28	29
11/12/69																				
6:30 am	-	-	-	-	-	22.8	-	1.5	.01	.01	-	-	-	-	-	-	-	-	-	-
11/13/69																				
6:30 am	52	-	77	83	86	20.8	-	4.1	1.0	.01	-	-	-	-	-	104	-	62	20	24
11/14/69																				
6:30 am	-	-	-	-	-	28.5	-	12.7	.05	.03	468	-	168	76	76	113	-	62	28	26

Remarks: 1, 2, 3, 4 are column effluents.

TABLE 33

Test 1

Carbon Columns Data

Date	Color % Trans.					Phenol (mg/l)					COD (mg/l)					TOC (mg/l)				
	TK	1	2	3	4	TK	1	2	3	4	TK	1	2	3	4	TK	1	2	3	4
11/15/69 4:00 am	31	-	70	78	83	38.0	-	25.5	.01	.01	-	-	-	-	-	-	-	-	-	-
11/16/69 3:45 am	34	-	51	63	68	37.3	-	30.0	.01	.01	322	-	237	93	85	103	-	85	44	43
11/17/69 6:45 am	48	-	70	74	77	31.4	-	30.0	.08	.01	-	-	-	-	-	127	-	84	59	42
11/20/69 6:30 am	29	-	-	60	66	31.0	-	-	.09	.01	315	-	-	111	86	141	-	-	35	33
11/23/69 7:05 am	71	-	91	91	93	19.5	-	-	1.3	.05	180	-	-	57	78	91	-	-	33	28
11/24/69 6:00 am	59	-	-	79	84	32.5	-	-	3.5	.02	300	-	-	150	100	114	-	-	53	46
11/25/69 6:30 am	66	-	-	85	85	-	-	-	-	-	233	-	-	150	116	88	-	-	49	41
11/26/69 10:00 pm	-	-	-	-	-	40.0	-	-	5.0	.01	-	-	-	-	-	132	-	-	43	42
11/28/69 8:00 am	-	-	-	-	-	25.0	-	-	7.1	.05	274	-	-	116	91	102	-	-	41	31
11/29/69 6:30 am	60	-	-	95	96	18.0	-	-	10.0	.01	166	-	-	75	63	-	-	-	-	-
12/1/69 8:20 am	77	-	-	96	97	32.0	-	-	14.8	.01	324	-	-	158	116	-	-	-	-	-
6:30 am	65	-	-	93	96	38.3	-	-	34.8	.01	394	-	-	271	156	-	-	-	-	-
12/2/69 6:30 am	74	-	-	94	98	281	-	-	32.5	.04	235	-	-	179	171	-	-	-	-	-
12/3/69 6:30 am	68	-	-	-	89	40	-	-	-	.12	-	-	-	-	-	-	-	-	-	-
12/4/69 6:30 am	-	-	-	-	-	27.5	-	-	-	.08	382	-	-	-	106	109	-	-	-	31
12/5/69 6:30 am	70	-	-	-	90	36.8	-	-	-	.26	366	-	-	-	74	112	-	-	-	24
12/6/69 4:30 pm	70	-	-	-	88	36.8	-	-	-	.26	363	-	-	-	198	110	-	-	-	34
12/7/69 3:30 pm	77	-	-	-	92	45.7	-	-	-	.57	330	-	-	-	140	-	-	-	-	-
12/8/69 6:25 am	61	-	-	-	86	24.1	-	-	-	1.7	331	-	-	-	98	-	-	-	-	-
12/9/69 10:30 pm	-	-	-	-	-	32.0	-	-	-	28.4	-	-	-	-	-	-	-	-	-	-
6:15 am	-	-	-	-	-	-	-	-	-	-	270	-	-	-	234	-	-	-	-	-
12/10/69 Final Sample	-	-	-	-	-	39.4	39.0	35.8	39.9	37.4	-	-	-	-	-	-	-	-	-	-

Remarks: 1, 2, 3, 4 are column effluents.

was selected because its breakthrough point occurred prior to that of phenol and COD. Figures 19-20 show a 5 ft wave front with a sharp breakthrough. The contact time required to contain the organics was 20 minutes based on a void bed volume. Based on experience by Calgon Corporation, a typical full scale treatment system for this wave front is generally sized for 60 minutes. This sizing factor provides an adequate safety factor and also results in the best relationship of operating cost vs initial cost. A 60 minute contact time would require design based on 8 ft^3 of carbon/gpm.

Test 2

This test was designed to test upflow procedures, to compare 8 x 30 mesh with 12 x 40 mesh activated carbon (Filtrasorb 400 - Calgon Corporation), and to test for the removal of cyanide through reactions of a catalyst (copper sulfate), pure oxygen and activated carbon.

In the upflow configuration, each column was filled to capacity with carbon except for a 1/4 - 1/2 inch area at the top of the column. The area at the top was filled with flexible material (steel wool) to provide good flow distribution. This system presented operational difficulties such as the entrainment of air, leakage of carbon, and plugging of the beds, which although solvable, did not warrant extra effort in the face of the project time table. Column flow was switched to down flow using only 3 columns for the remainder of the test and operated in the same manner as in the initial test except that a flow rate of 1/2 gpm through the columns was utilized during the latter part of the test. The treated effluent from the carbon columns was collected in an additional holding tank. This treated waste water was pumped at 20 psi through a small mixing cylinder where additions of copper sulfate catalyst and commercially pure oxygen were made. The oxygen was actually injected into the water stream just ahead of this tank at a pressure 4 to 5 lbs above the water pressure. The catalyst consisted of a 12,000 mg/l solution of commercial grade copper sulfate in plant water. The flow of 1/2 gpm passed to two steel columns for cyanide destruction, each containing 60 pounds of activated carbon. Initially the columns were operated in series; however, the 2nd column did not receive sufficient catalyst and the units were switched to parallel operation with a flow rate of 1/4 gpm through each column. Catalyst concentrations four times the concentration of the cyanide (influent concentration) and an effluent containing no less than 4 mg/l dissolved oxygen were the operating criteria used during the test. Chemical reactions occurring during the treatment of the cyanide wastes are:

Total Oxidation:

$$2\ CN + 2.5\ O_2 + H_2O = 2\ CO_2 + N_2 + 2OH$$

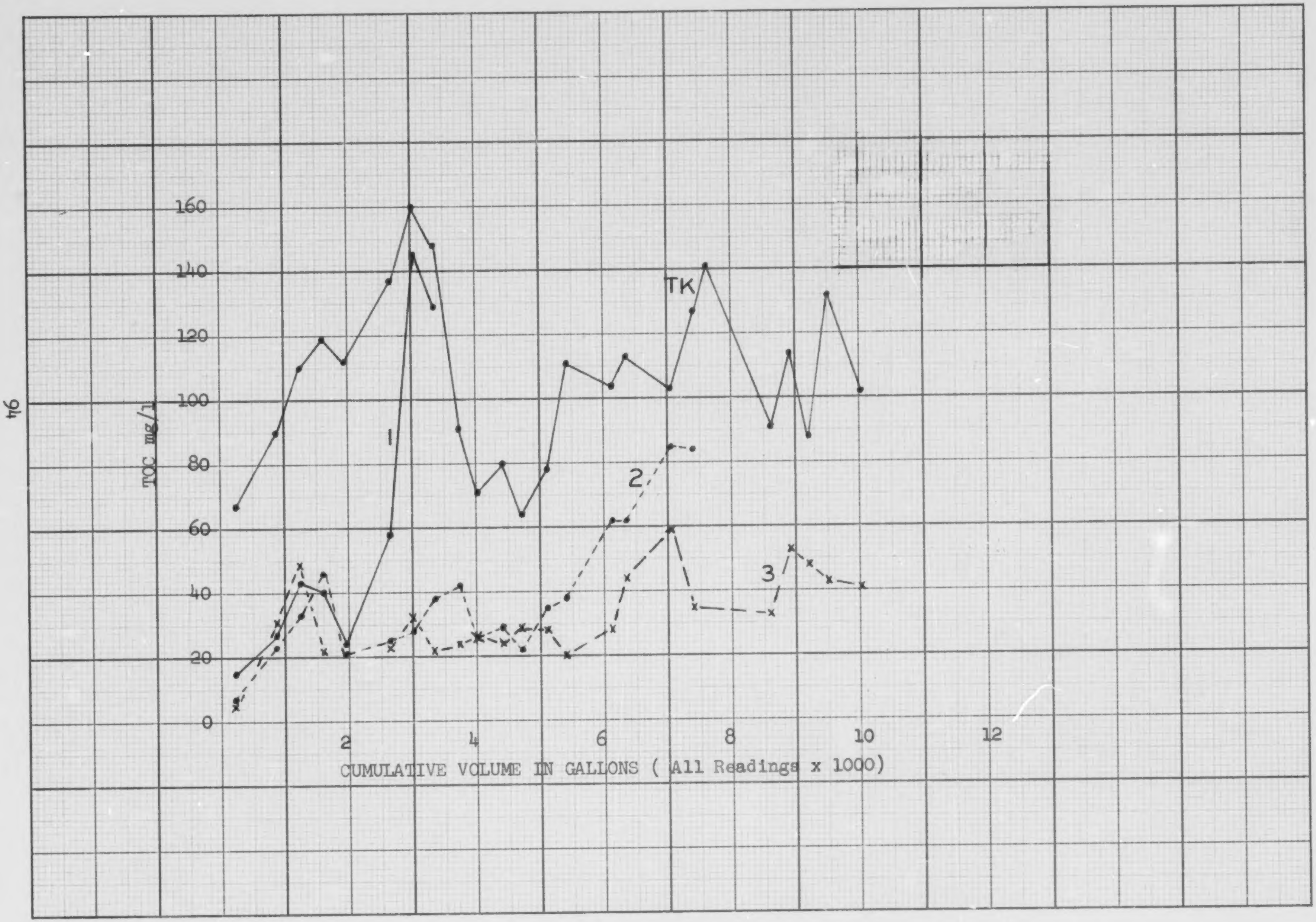


Figure 19. TOC BREAKTHROUGH, TEST 1

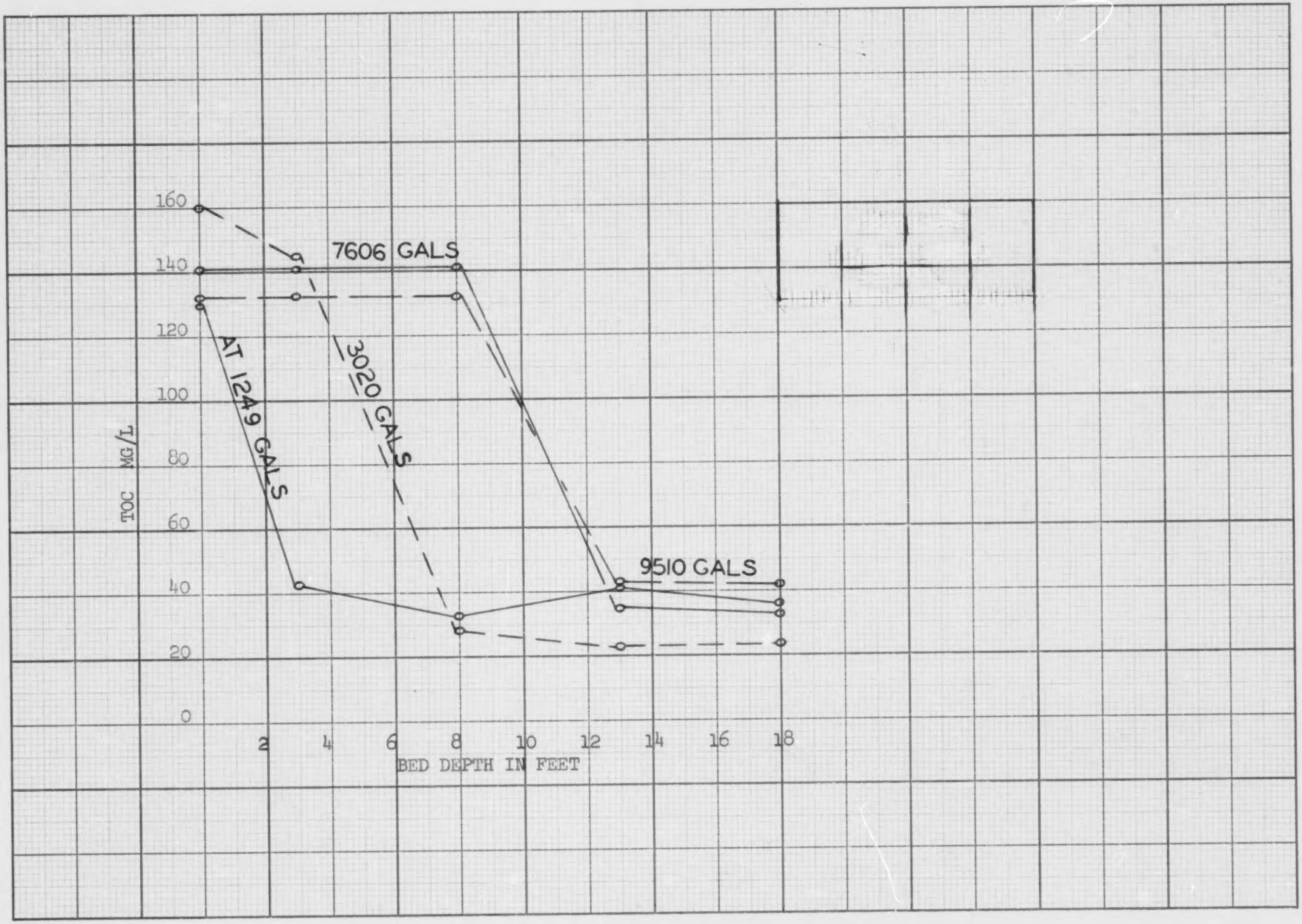


Figure 20. TOC WAVE FRONT, TEST 1

Partial Oxidation:
$2\ CN + O_2 = 2\ CNO$

Destruction of the Cyanate:
$CNO + 2\ H_2O = NH4 + CO_3$

The second test, primarily designed to obtain cyanide destruction data, was not run long enough to provide additional breakthrough data on the carbon adsorption columns. However, the data obtained (Tables 34-35) confirmed the results obtained in the previous test. No significant difference was observed in the adsorption capabilities of Filtrasorb 300 and Filtrasorb 400.

Table 34 summarizes the data from the cyanide removal columns. The influent cyanide concentration ranged from 15 to 48 mg/l while the maximum effluent concentration was 0.30 mg/l. The detention time in the columns was 75 minutes. For an actual full scale system, the detention time may be reduced to one-half that used in the pilot plant, or 37.5 minutes. A 37.5 minute detention time would require a system design based on 5 ft^3 of Filtrasorb/gpm.

Test 3

The third test was conducted to find the adsorption wave front and Filtrasorb organic capacity at high dissolved organic levels using Filtrasorb 400 (12 X 40) mesh activated carbon. The test procedures were identical to the previous test; that is, downflow through a fixed-bed-in-series configuration followed by the cyanide removal units operated in parallel. Flow rates were maintained at 1/4 gpm through the carbon adsorption columns. In order to provide a flow of 1/4 gpm through each of the cyanide removal columns, the adsorption columns were run for a period before the cyanide removal columns were actuated. Only one cyanide removal column was operated for a short period. The high concentrations of organics resulted in rapid breakthrough in the carbon columns. See Tables 36-38. Figure 25 shows the breakthrough curves for T.O.C. From these curves the dosage is approximately 31 lbs per 1000 gallons. The narrow wave front curves on Figure 22 show a 9 ft wave front at 0.25 gpm or a detention time of 36 minutes, based on void volume. Full scale units should be designed for 108 minutes detention time, requiring 14.5 ft^3 Filtrasorb/gpm. Because the contaminant concentration rises rapidly after the breakthrough, Calgon Corporation recommended a design safety factor of 3.

The cyanide destruction columns showed the results of rapid breakthrough of organics in the columns. Good reduction was obtained, however, maximum effluent levels gradually increased during the test. See Table 37.

TABLE 34

Test 2

Cyanide Removal Columns C-7A (12x40 Mesh)

Date	Type of Sample	pH Influent	pH Final Effluent	CYANIDE Influent mg/l	CYANIDE Final Effluent mg/l	COPPER CATALYST Concentration Influent mg/l	COPPER CATALYST Concentration Effluent mg/l	Effluent Dissolved Oxygen mg/l
12-29-69	Composite	--	--	34	0.1	--	--	--
1-14-70	Composite	8.6	9.1	20	0.05	137	0.20	--
1-15-70	Composite	11.4	11.9	42	0.04	69	0.15	--
1-16-70	Composite	7.9	7.0	41	0.23	94	0.15	20
1-17-70	Grab	7.4	6.4	48	0.04	56	0.15	--
1-18-70	Composite	7.4	7.5	15	0.25	60	0.20	12
1-19-70	Composite	7.2	7.8	15	0.30	2.2	1	--
1-20-70	Composite	8.6	8.0	18	0.20	29	5.6	12
1-21-70	Composite	9.1	9.1	15	0.07	30	2.7	13
1-22-70	Composite	9.2	9.0	16	0.13	25	5.8	1

The total flow rate through the system was maintained at 0.25 gallons per minute.

TABLE 35

Test 2

Carbon Columns 7A (12x40 Mesh)

Date	Phenol (mg/l)				Color % Trans.				COD (mg/l)				TOC (mg/l)			
	TK	1	2	3	TK	1	2	3	TK	1	2	3	TK	1	2	3
Upflow Units																
12/26/69 2:30 pm Grab (Plugged Up)	19.2	.01	.01	.01	-	-	-	-	-	-	-	-	-	-	-	-
Downflow Units																
1/13/70 2:30 pm Grab																
Composite 10:30 pm	32.5	.08	.03	.05	27	73	96	98	592	137	29	33	346	60	36	42
6:30 am	32.7	.08	.01	.01	39	80	77	87	383	83	91	96	78	37	40	48
1/14/70 Composites 2:30 pm 10:30 pm 6:30 am	33.0	.14	.01	.01	1	13	29	54	491	203	174	100	130	98	41	53
1/15/70 Composites 2:30 pm 10:30 pm 6:30 am	31.1	.02	.01	.01	2	2	50	65	681	456	120	124	131	47	45	48
1/16/70 Composites 2:30 pm 10:30 pm 6:30 am	40.0	.23	.01	.01	42	76	81	86	484	134	147	124	213	58	53	56
1/17/70 6:30 am Grab	21.2	.85	.01	.01	70	78	90	95	237	102	76	59	138	57	42	42
1/18/70 Composites 8:00 pm 7:00 am	37.5	3.3	.16	.01	61	83	86	92	336	34	67	76	154	61	47	49
1/19/70 Composites 10:30 pm 6:30 am	21.1	8.0	.03	.01	43	72	81	86	395	58	59	35	255	52	33	35
1/20/70 Composites 2:30 pm 10:30 pm 6:30 am	38.5	39.5	.18	.01	117	70	78	78	532	316	125	125	275	164	63	64
1/21/70 Composites 2:30 pm 10:30 pm 6:30 am	33.8	34.0	.90	.01	29	64	75	-	655	438	186	131	244	226	57	53
1/22/70 Composites 2:30 pm 10:30 pm 6:30 am	23.4	39.6	6.0	.01	50	78	85	89	562	312	115	167	280	156	68	49

Remarks: 1, 2, 3, 4 are column effluents.

TABLE 36

Test 3 Activated Carbon

Ammonia Flushing Liquor

Date	Hours Operated	Flow Rate Gallons Per Minute Carbon Columns TOTAL	Cyanide 1	Columns 2	Gallons Per Day Carbon Columns TOTAL	Cyanide 1	Columns 2	Cumulative Gallons Total Carbon Columns TOTAL	Cyanide 1	Columns 2
1-28-70	19	.26	-	-	295	-	-	295	-	-
1-29-70	22.5	.25	.20	-	342	231	-	637	231	-
1-30-70	23	.30	.20	-	416	277	-	1053	508	-
2-2-70	18	.26	.18	.24	285	253	255	1338	761	255
2-3-70	21	.27	.27	.22	341	288	278	1679	1049	533
2-4-70	4	.25	.24	.26	61	57	63	1740	1106	596

TABLE 37

Test 3

Cyanide Removal Columns - Ammonia Flushing Liquor

Date	Color % Trans			Cyanide (mg/l)			COD (mg/l)			TOC (mg/l)		
	TK	1	2	TK	1	2	TK	1	2	TK	1	2
1/29/70 Composites 8:00 am 11:00 pm	70	98	--	20	1.5	--	302	42	--	317	271	--
Composites 4:00 am 6:00 am	66	100	--	16.8	1.3	--	463	78	--	352	336	--
1/30/70 2:30 pm Grab	66	100	--	14	13.5	--	326	363	--	--	--	--
2/2/70												
2:30 pm Grab	24	99	99	10	.72	.07	566	193	--	--	--	--
6:00 am Grab	1	94	97	7.8	.9	1.0	1587	245	22	--	--	--
2/3/70												
2:30 pm Grab	1	98	100	10.5	1.5	1.4	2100	970	875	--	--	--
6:00 am Grab	1	99	99	10.6	2.2	2.5	--	500	436	--	--	--
2/4/70 10:00 am Grab	--	--	--	5.9	2.1	1.3	--	--	--	--	--	--

Remarks: 1, 2, 3, 4 are column effluents.

TABLE 38

Test 3

CARBON COLUMNS DATA - AMMONIA FLUSHING LIQUOR

Date	Color % Trans.					Phenol (mg/1)					COD (mg/1)					TOC (mg/1)				
	TK	1	2	3	4	TK	1	2	3	4	TK	1	2	3	4	TK	1	2	3	4
1/28/70 Comp.	1	41	92	98	100	682	91	.01	.01	.01	3387	900	332	300	288	3040	960	150	165	188
1/29/70 12:30 Grab	-	-	-	-	-	691	643	.42	-	-	-	-	-	-	-	-	-	-	-	-
9:00-2:30 Comp. pm	1	1	87	100	100	700	588	158	.01	.12	3278	2830	343	408	375	2848	2152	185	216	193
8:00-11:00 pm Comp.	3	7	70	86	99	760	1000	.8	.01	.01	4077	3370	1206	957	790	3596	3320	462	406	394
4:00-6:00 am Comp.	3	21	38	76	94	762	882	423	.08	.01	3993	3244	1830	541	426	3532	3248	1304	394	410
1/30/70 9:10 am After B	-	-	-	-	-	-	550	-	-	-	-	2621	-	-	-	-	-	-	-	-
2:30 pm Grab	5	6	6	68	98	740	960	1100	13	.01	3619	3619	3577	631	505	-	-	-	-	-
6:00 am Grab	5	7	50	69	84	760	834	920	846	.08	4227	3495	3295	2743	1219	-	-	-	-	-
2/2/70 2:30 pm Grab	4	23	39	43	56	678	582	992	1132	15.9	4374	3602	3430	3430	589	2051	1662	1641	1484	319
6:00 am Grab	1	1	2	10	21	824	854	890	910	734	4459	3730	3378	3130	2444	1593	1765	1561	1363	1232
2/3/70 2:30 pm Grab	1	7	28	43	58	810	910	1014	1120	-	4685	4120	3990	3860	-	1966	1850	1718	1691	910
6:00 am Grab	1	1	5	15	49	811	824	892	903	978	3830	3830	3574	3574	2894	2156	2060	1877	1606	1540
2/4/70 10:00 am Grab	-	-	-	-	-	815	830	870	890	985	-	-	-	-	-	-	-	-	-	-

Remarks: 1, 2, 3, 4 are column effluents.

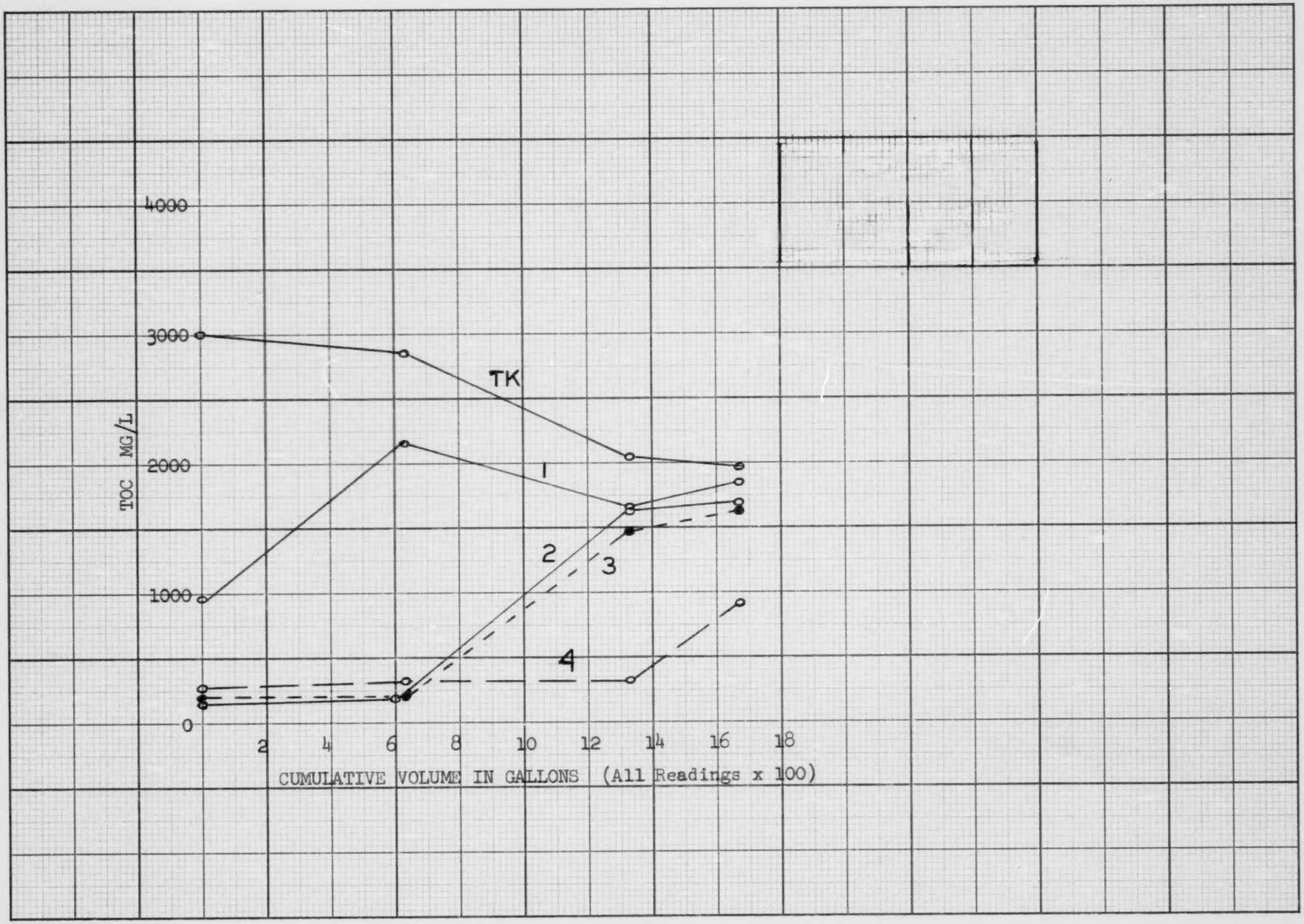


Figure 21. TOC BREAKTHROUGH, TEST 3

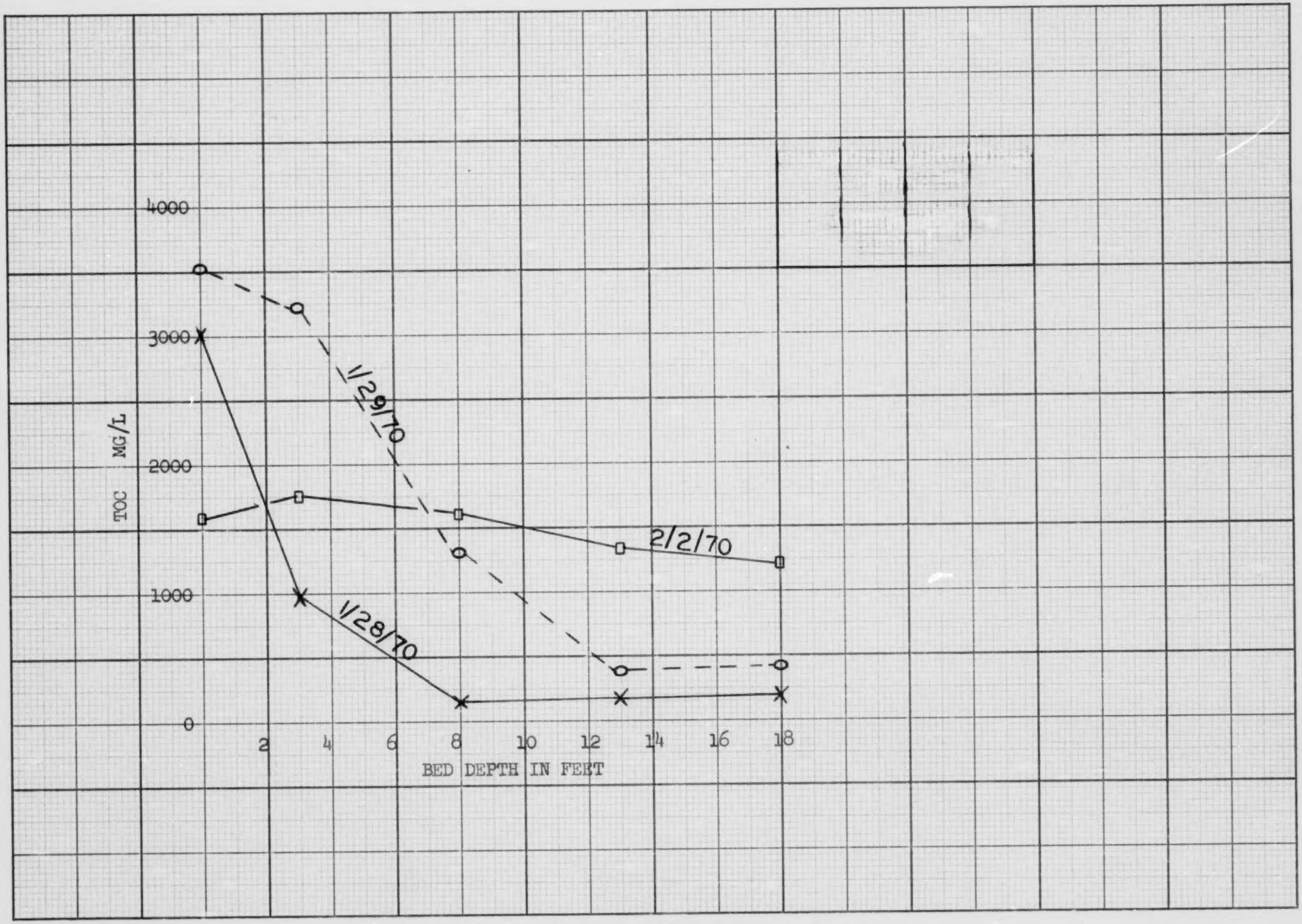


Figure 22. TOC WAVE FRONT, TEST 3

Spot checks were made during all three tests to see if there was ammonia removal. There was none. Tests on cyanide removal in the standard carbon adsorption columns were made, but no reductions were noted.

Temperature fluctuations in the waste feed from 2°C to 38°C did not affect the carbon columns nor the cyanide destruction column operation.

Since a lower pH generally enhances adsorption, an isotherm was run on samples from sampling point C-7A at pH values of 8.9 and 6.8. The samples contained T.O.C. concentrations of 500 mg/l. Filtrasorb dosages at a pH of 6.8 were 39% lower than those for a pH of 8.9. This lower dosage represents an initial savings in activated carbon of about $157,000 and a reduction in annual operating costs of $26,000, based on 5% regeneration loss and 80% T.O.C. removal.

The operating costs for using Filtrasorb were determined by the previously described flow-through data as well as the manufacturer's product specifications, as follows:

pH Avg.	T.O.C. Mg/1	Filtrasorb Dosage Lbs./1000 Gallons	Operating Cost ¢ 1000 Gallons
8.9	64-160	4.5	8
8.9	500	8.2	14.7
8.9	1932-3532	31	55.5

The carbon cost including regeneration is 1.79¢/pound for a coke plant flow of 2000 gpm, a T.O.C. concentration of 500 mg/l, and a contact time of 60 minutes. The following are supporting figures:

	Cost/Pound
Make-up carbon - assuming a 5% loss of carbon per regeneration and based on 6 regenerations per year.	
8,640,000 Lbs. of carbon regenerated/year 432,000 Lbs. of make-up carbon needed/year Cost of carbon 28¢/pound	1.4¢/Lb.
Operating Labor - Operator attention is estimated at six hours per day for the regeneration system. Assume $6 per hour, including fringe benefits.	.152¢/Lb.

	Cost/Pound
Fuel Gas - For the regenerating furnace approximately 4,000 BTU is required per pound of carbon regenerated. The cost to regenerate 8,640,000 pounds of carbon, using 1,000 BTU/ cu. ft. gas at 30¢/MCF, is .12¢/pound of carbon	.12¢/Lb.
Electric Power Regeneration System - 7.5 KW operating load, operating 365 days per year: 66,000 KWH/yr. Feed Pump - 25 KW operating load, operating 24 hours a day, 365 days a year: 220,000 KWH/yr. Backwash System - 9 KW operating load, operating one hour per day, 365 days per year: 3,350 KWH/yr. Effluent Pump - 4.5 KW operating load, operating 24 hours a day, 365 days a year: 39,400 KWH/yr. Utility Pump - 6.6 KW operating load, operating 24 hours a day, 365 days a year: 58,000 KWH/yr. A total of 387,000 KWH/yr. at 1¢ per KWH costs .045¢/Pound of carbon regenerated.	.045¢/Lb
Steam - One pound of steam is used for each pound of carbon regenerated. Steam costs are 53¢/1000 pounds of steam.	.053¢/Lb
Maintenance - The carbon system should require very little maintenance. Approximately $3,000 per year is estimated.	.035¢/Lb
Total Operating Expense	1.79¢/Lb

Annual Operating Cost Carbon Columns

Make-up Carbon	121,960
Operating Labor	13,140
Fuel Gas	10,368
Electrical Power	3,890
Steam	4,580
Maintenance	3,000
	$154,938/year

Capital costs for a 2000 gpm activated carbon plant, estimated by Calgon Corporation to be in the $700,000 - $800,000 range, were obtained from Figure 13. This curve has been prepared based on engineering estimates by Calgon Corporation and its validity has been confirmed by the cost of several plants either completed or under construction. The above estimated cost is of course only an order of magnitude estimate and actual cost would be defined only by a detailed engineering study. The design and operating cost estimates are based on a moving bed counter flow design and a carbon regeneration by a multiple hearth, gas-fired furnace. With this type of system fresh carbon would always be available at the effluent section for final polishing. Results similar to the first days operation of the pilot unit could be expected.

An important advantage of the activated carbon treatment method is its ability to handle large increases in organic loading without a large increase in system detention time. For example, the average TOC loadings in test 3 were 334 mg/l as compared to 103 mg/l in test 1. However, a full scale treatment system requires a detention time of 108 minutes at the higher TOC loading as compared to 60 minutes at the lower TOC loading. Surge increases in loading which are expected from the coke plant by-product operations could be handled in a properly designed system and would not bring about a prolonged system upset. Another advantage of the carbon system is that minimum pretreatment and storage capacity are required. Although Calgon Corporation consultants feel confident that the testing program provided sufficient data for engineering design of a full scale cyanide destruction facility, it is the opinion of CF&I personnel that the engineering study should include additional testing of the cyanide destruction column to determine operating procedures and parameters for five to six months continuous operation. Among these are catalyst fouling and frequency and methods of carbon regeneration as a function of prolonged adsorption of materials passing through the columns.

Capital and operating costs for the cyanide destruction unit are not clearly defined because of its recent development and consequent lack of data from full scale treatment plants. Calgon's estimate of the capital cost for a 2 MGD plant is $300,000 - $400,000. Estimated operating costs are based on complete removal of cyanides and includes make-up catalyst and oxygen. After initially conditioning the system with catalyst, the estimated operating cost would be $0.030/1000 gallons treated. The oxygen requirements which were incorporated into this cost were figured at $.60/1000 ft^3. For CF&I's 2000 gpm waste stream, the annual cost of oxygen and catalyst would be $31,645.

Adsorption Capacity of CF&I Coke

A sample of coke was sent to Nalco's Ion Exchange Laboratory for evaluation of its capacity to adsorb phenols and cyanides. A 55 gallon drum containing ammonia flushing liquor was shipped to Chicago for use in this study.

Analysis of Flushing Liquor from CF&I

a.	Ammonia as NH_3	1400 mg/l
b.	Cyanide as CN	48 mg/l
c.	Phenols as C_6H_5OH	640 mg/l
d.	TOC as C	1600 mg/l
e.	COD as oxygen demand	4500 mg/l
f.	Total Hardness as $CaCO_3$	27 mg/l
g.	Sodium as Na_2O	11 mg/l
h.	pH	9.5

Apparatus and Equipment

a. 1" ID Lucite column with Saran screen bottom
b. Eastern Industries centrifugal pump.
c. Beckman pH meter
d. B&L "Spectronic 20" spectrophotometer
e. Schute & Koerting flowmeters

Preparation of Feed Solution

The flushing liquor received in a 55 gallon steel drum was strained through glass wool inserted in a funnel to remove particles of dirt or coal. These particles were removed to avoid problems in column operations.

Preparation of Coke

The sample of coke was screened to include particle sizes of - 12 to 50 mesh. Lumps were broken up by forcing them through a 12 mesh screen. The screened material was transferred to the 1" Lucite column and back-washed with demineralized water to remove color and fines.

Column Operation of Coke

The flushing liquor was pumped through the 1" ID Lucite column at a flowrate of 2 gals./ft^3 coke/min. Effluent samples were collected at regular intervals for NH_3, CN, Phenol, TOC and pH analyses.

Effluent Analysis

	Flushing Liquor	Coke Effluent	% Removal
Ammonia as NH_3	1400 mg/l	1300 mg/l	7.1
Cyanide as CN	48 mg/l	37-43 mg/l	10-24
Phenols as Phenol	640 mg/l	600-720 mg/l	0
Total Organic Carbon	1600 mg/l	1800 mg/l	0
pH	9.5	9.5	

There was no color reduction of the flushing liquor in the effluent. The above analyses were obtained from an effluent volume of 16 bed volumes during one run.

Results & Conclusion

The results of this screening test indicate that coke does not show promise as an adsorbent for the coke plant waste water contaminants.

Adsorption Test CF&I Fine Coal

A sample of fine coal was tested in the laboratory to determine its adsorptive capacity. Specific tests included the determination of Iodine Number, Molasses Number, and the adsorption isotherm in comparison with activated carbon.

The Iodine Number is the number of milligrams of iodine adsorbed per gram of carbon when in equilibrium (under specified conditions) with a 0.02N Iodine solution. The Iodine Number is an approximate measure of the adsorptive affinity for small molecules.

The Molasses number is a measure of the adsorptive affinity for color bodies in a specified molasses solution, as compared to a standard carbon. The adsorbance of the filtrates are determined on an electrophotometer and the molasses number of the unknown is calculated from the ratio of the adsorbance values of the sample and the standard carbon. In this case the affinity is for large molecules, such as color bodies.

An adsorption isotherm is the ratio of the amount of substance adsorbed to the amount remaining in solution, expressed as percent.

	CF&I Coal	Filtrasorb 300
Molasses Number	160	200
Iodine Number	286	975
Ash (percent)	10.8	5

Adsorption Isotherm on 1000 mg/l phenol solution

a.	CF&I Coal	less than 0.1% by weight pickup
b.	Filtrasorb 300	15% by weight pickup

From the adsorptive capacity indicated by these screening tests, the use of coal as an adsorption material for treating coke plant wastes was not given further consideration. It had been contemplated that large beds of coal might be suitable for treatment purposes with the spent coal being periodically charged into the coke ovens. This procedure might have resulted in problems associated with recycling the adsorbed contaminants which would have required further evaluation.

CHEMICAL TREATMENT AND CLARIFICATION

Laboratory Investigation of Lime Treatment of Ammonia Flushing Liquor

The ammonia which is formed during the coking operations is found in the gas produced and in the water vapors condensed by the flushing liquor. Ammonia is present in the flushing liquor both as free and fixed ammonia. Free ammonia can be removed by stripping with an inert gas such as air. Addition of heat decreases the solubility of the free ammonia resulting in more complete removal. The fixed ammonia requires the addition of strong alkali for conversion of fixed ammonia to free ammonia prior to removal by stripping. Salts such as ammonium chloride, sulfate, thiocyanate, ferro cyanide, etc., are considered as fixed ammonia compounds.

Possibilities for removal of ammonia from the flushing liquor at CF&I include:

1. Steam stripping of flushing liquor for removal of free ammonia.

2. Removal of both free and fixed ammonia in the ammonia still by operation of both the free ammonia leg and the lime leg of the still.

Item 1 above is most easily accomplished. Complete removal, however, as indicated under Item 2, is much more troublesome inasmuch as the lime leg part of the still requires considerably more attention and maintenance.

Elimination of the lime leg, or external treatment of the flushing liquor with lime before entrance to the still, would be an improvement over present procedures.

The work described here was conducted in order to establish the most obvious parameters of the lime treatment of the flushing liquor.

In actual plant application the stripped ammonia would be directed back into the gas stream to be removed by contacting with dilute sulphuric acid in the saturators the same as the rest of the ammonia in the gas. The resultant ammonium sulphate is utilized in the production of fertilizer. Because of marketing conditions, however, this operation is normally conducted at a loss based only on acid costs. The ammonia processing is therefore conducted essentially to prepare the coke gas for beneficial use and to prevent water pollution.

Purpose:

1. To establish lime requirements for pH adjustment (lime-pH titration curves).
2. Determine degree of conversion of fixed ammonia to free ammonia with increasing values of pH.
3. Investigate changes in cyanide, phenol, and COD levels after addition of lime.

Procedures:

The lime-pH titration curves were obtained by the addition of a lime slurry of known concentration to 500 ml samples of ammonia flushing liquor. The pH was determined on a Beckman Zeromatic, line operated pH meter. A 100 gram per liter lime slurry was used. Slurry additions were made by means of a graduated pipette with mixing between additions.

The conversion of fixed ammonia to free ammonia was established by adding lime to a 500 ml sample until a desired pH was obtained. The sample was then boiled for exactly 10 minutes, and allowed to cool. A distillation analysis for ammonia nitrogen was run immediately after cooling.

The study of cyanide, phenol, and COD concentration as a function of pH was made using 3 liter samples. The pH was increased to five different levels in the range of pH 8.4 - 8.6 (that of the raw sample) to pH 11.2. Commercially available hydrated lime was used.

Sludge volume was determined by transferring 1 liter to a graduated cylinder and settling for 30 minutes.

Some samples were treated with spent pickle liquor preceeding adjustment to pH 10.5. The purpose of the addition of pickle acid was to determine if cyanide could be precipitated as a complexed iron salt.

Results:

Lime-pH titration curves are shown on Figures 23-25. These results were obtained on three different sampling days.

It is apparent that the pH increases rapidly once the level of 9.5 has been reached. This requires approximately 10 lbs. of lime per

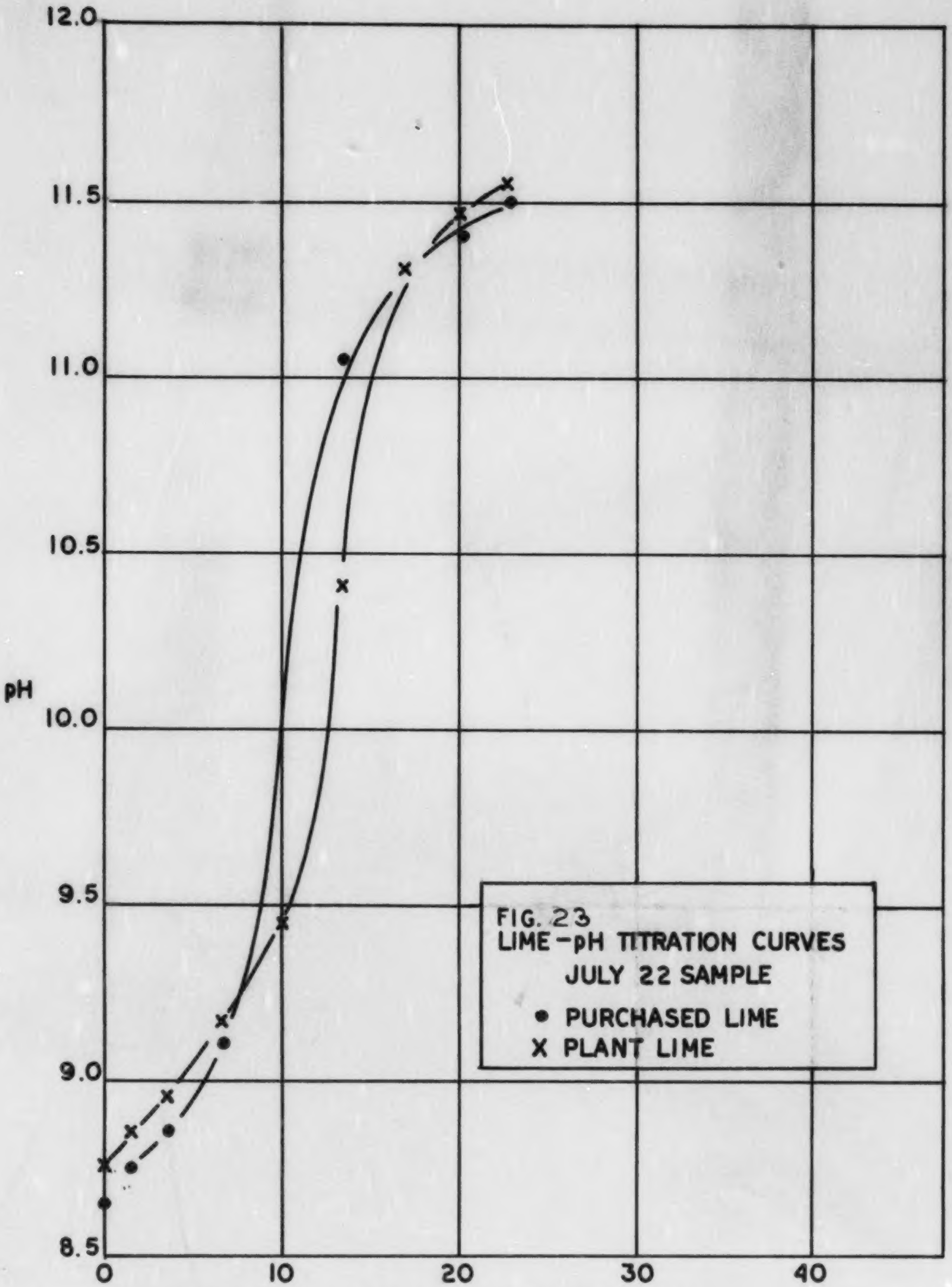


FIG. 23
LIME–pH TITRATION CURVES
JULY 22 SAMPLE

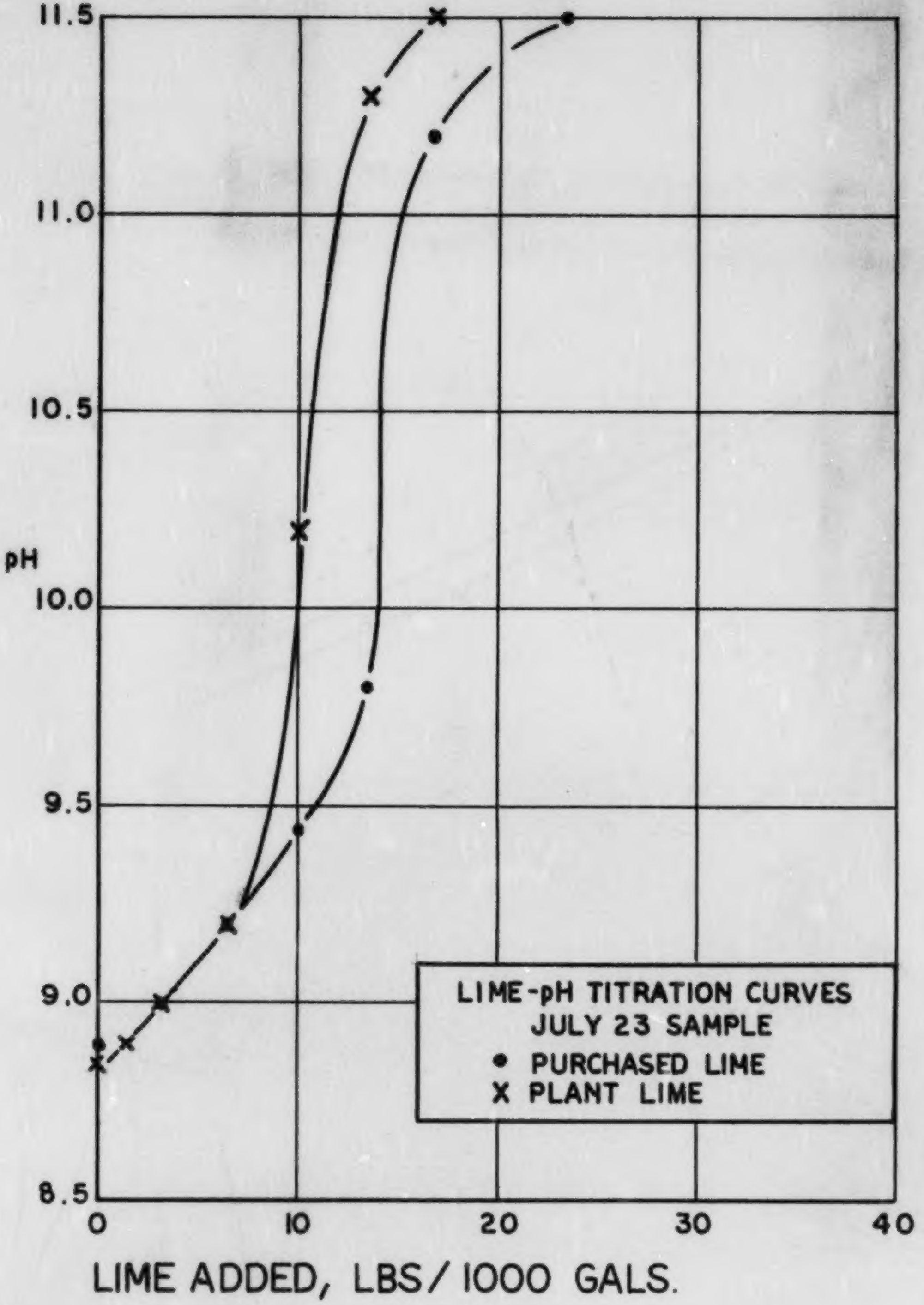

11.5
11.0
10.5
10.0
9.5
9.0
8.5
pH
0
10
20
30
40
LIME ADDED, LBS/1000 GALS.
LIME-pH TITRATION CURVES
JULY 23 SAMPLE
● PURCHASED LIME
X PLANT LIME

FIG. 24

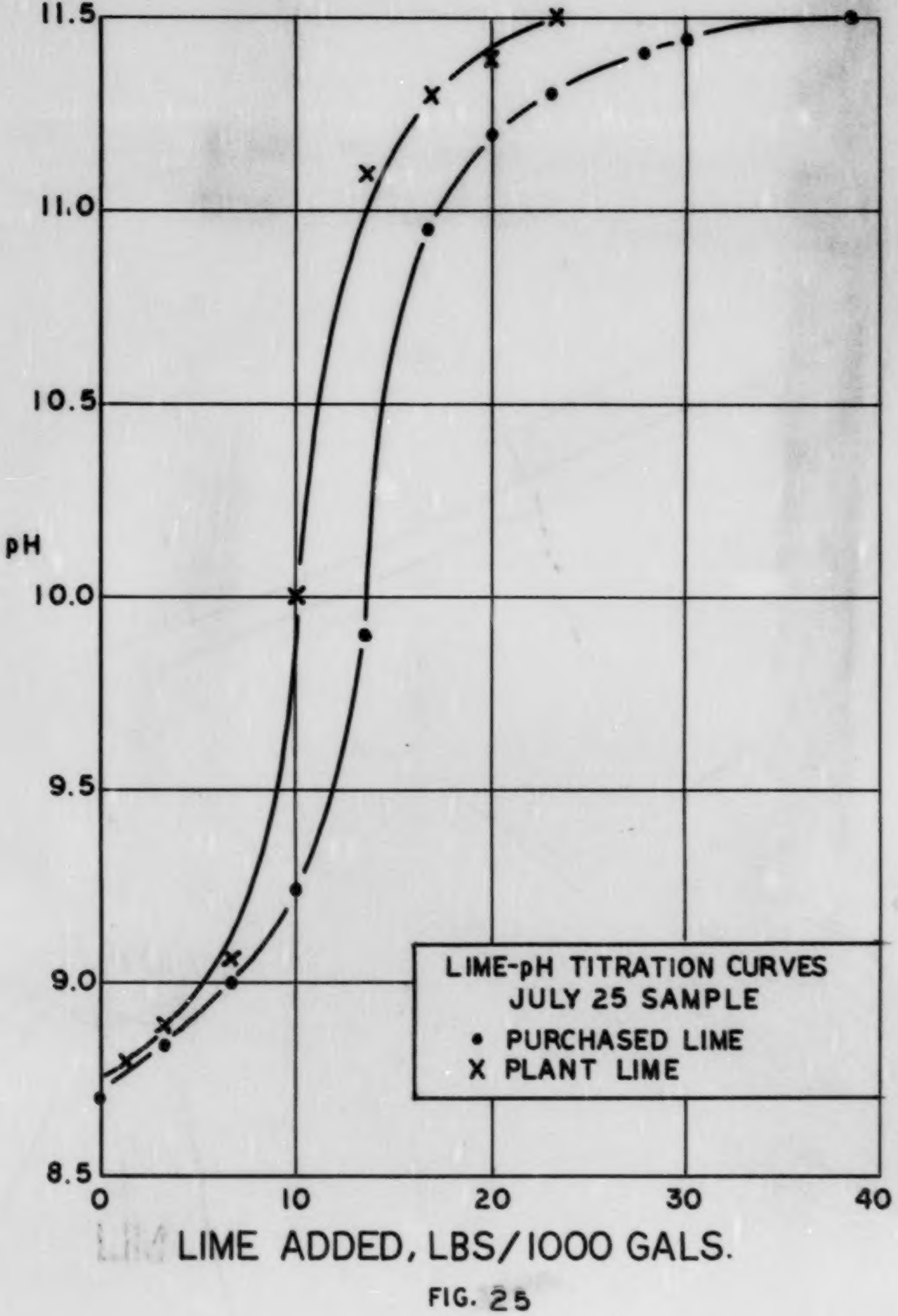

LIME-pH TITRATION CURVES
JULY 25 SAMPLE
• PURCHASED LIME
X PLANT LIME
pH
11.5
11.0
10.5
10.0
9.5
9.0
8.5
0
10
20
30
40
LIME ADDED, LBS/1000 GALS.

FIG. 25

1,000 gallons of flushing liquor. The quantity of lime required to reach pH 11.5 varied somewhat with the sample and the type of lime used. The plant lime, as produced by CF&I, was obtained directly from the kiln and was expected to be essentially calcium oxide. The hydrated lime purchased outside of CF&I was obtained from the plant lime soda softener operation in the boiler house.

According to the data shown in Figures 23-25 the quantity of lime required to raise the pH of the flushing liquor to 11.2 may vary between 15.5 and 20 lbs. of purchased lime per 1,000 gallons flushing liquor. The quantity of plant lime needed to raise the pH to 11.2 varied between 13 and 16.5 lbs. of lime per 1,000 gallons flushing liquor. The April, 1970, cost of hydrated lime is .0155¢/lb. as compared to the plant lime of .0065¢/lb. Although a higher dosage is required, the use of CF&I quick lime is indicated.

The characteristics of lime-treated ammonia flushing liquor collected on July 22, 23, and 25 are shown in Table 39. This table shows substantial reduction in ammonia at pH levels above 10.4. It should be noted that these reductions were accomplished by conversion of the fixed ammonia to free ammonia followed by removal by boiling for 10 minutes. The values shown in the Tables are levels of ammonia nitrogen remaining after that treatment.

Some reduction in cyanide was noted with the addition of the first increment of lime. Thereafter, no further reduction was obtained. The July 25, sample showed cyanide reduction as a result of the first increment of lime. There was no explanation available for this result. This same sample showed very poor reduction in ammonia at higher pH levels as compared to the previous tests.

Sludge concentrations increased with lime additions as would be expected. Some of the solids increase in the July 22-25 tests is attributable to the precipitation of calcium and magnesium hardness as calcium carbonate and magnesium hydroxide.

As shown in Table 40, no significant reduction in cyanides could be obtained as a result of addition of spent pickle acid. Sludge volumes increased with increased volumes of acid added, mainly because of larger quantities of iron hydroxide following pH adjustment.

There was no conclusive evidence that addition of pickle acid improved the removal of any of the contaminants from the flushing liquor.

TABLE 39

LIME TREATMENT OF AMMONIA FLUSHING LIQUOR

July 22 Sample

Test No.	Lime Added lbs/1000 gal.	pH	NH_3* mg/1N	CN mg/1	Phenol C_6H_5OH mg/1	COD mg/1	Sludge Vol. ** gals/1000
Untreated	0	8.45	1624	48.4	670	5217	-
1	13	9.2	476	36	568	5050	10
2	16.7	9.3	484	36	550	-	10
3	21.7	9.7	401	39.2	430	5218	12
4	27.9	10.75	34	-	430	5134	20
5	46.7	11.35	68	-	418	5050	30
July 23 Sample							
Untreated	0	8.4	404	46	710	5258	-
1	14	9.2	370	37.6	670	5198	10
2	21	9.5	276	38.4	670	5188	12
3	25.3	9.9	-	37.6	632	5092	15
4	28	10.4	72	38.4	630	5153	16
5	33.7	11.2	81	38	620	5195	27
July 25 Sample							
Untreated	0	8.55	1632	48	810	5090	-
1	15.4	9.0	365	14.8	750	4990	10
2	22.4	9.35	268	-	735	-	16
3	26.5	9.8	289	15.4	691	4870	17
4	29.3	10.25	178	21	772	-	18
5	37.7	11.1	-	24.8	787	4910	32

Remarks: Adjust of pH with lime, followed by 30 minute settling
*Total ammonia determined after boiling for 10 minute
**Sludge vol. after 30 minutes settling

TABLE 40

Results of Lime Treatment of Ammonia Flushing Liquor after Addition of Pickle Acid

	RAW SAMPLE ANALYSIS					TREATMENT		TREATED SAMPLES				
Sample Date	pH	NH_3 mg/l	CN mg/l	COD mg/l	Phenol C_6H_5OH mg/l	Pickle Acid gals/1000 gals	Lime Added lbs/1000 gals	pH	CN mg/l	Phenol C_6H_5OH mg/l	COD mg/l	Sludge Vol.* gals/1000 gals
7-23	8.4	404	46	5258	710	.166	29.3	10.6	43.6	620	5474	--
						.33	30.15	10.5	43.6	595	5526	--
						.50	30.15	10.5	45.6	620	5203	--
7-24	8.5	1632	46.5	5202	670	.166	31	10.5	35.8	649	5017	20
						.33	30.15	10.5	38.4	332	4506	22
						.50	29.6	10.5	43.6	330	4990	28
7-25	8.5	1632	48	5090	810	.166	30.6	10.5	22.4	567	4780	17
						.33	30.3	10.5	36.9	586	4820	20
						.50	30.6	10.5	42	-	4990	22

Remarks: * Sludge Vol. after 30 min. settling.

Clarification

The total suspended solids (TSS) concentration in the total coke plant effluent, (Sampling Point C-7A) varies considerably. Extremely high solids loadings occurring during the operation of the coal washery while equipment at the washery is drained for maintenance and repairs will be eliminated as pollution control equipment is installed for that area. When the coal washery is down, the total suspended solids are in the range 80-300 mg/l. These solids are primarily coke breeze from the quench stations. Primary clarification is needed to reduce these solids to a level where final clarification can be accomplished in an activated carbon unit or, if biologically treated, in the final clarifier in that system. The reduction of suspended solids to the range 20-40 mg/l is considered a prerequisite for most final treatment methods.

Processes involving gravity sedimentation and/or filtration are adversely affected by grease or oil. These contaminants change the settling characteristics of the suspended solids resulting in less efficient solids removal by gravity sedimentation. Processes using filtration for solids removal become clogged by oil and grease and normal backwashing procedures do not adequately correct the problem. For this reason methods were considered for removing these substances prior to suspended solids clarification.

Essentially all oil discharges sampled at Point C-7A originate at the agitators in the benzol plant, the light oil separators, and the cresote plant. The intercepter sump downstream from C-7B is equipped with a skimming baffle to retain and remove free oil. Periodically caustic is discharged which emulsifies the accumulated oil at this point. At the present time, practically no oil is being removed mechanically. The elimination of caustic discharges and the periodic removal of accumulated free oil from the intercepter sump will eliminate practically all of the oil and other flotable material which are now discharged at C-7A.

A study was made of suspended solids removal by gravity sedimentation and filtration. Figures 26 and 27 show results of settling tests conducted in 2000 ml beakers on samples taken from C-7A with and without the coal washery operating. The tests were run by allowing the samples to stand undistrubed for a given period of time between withdrawal of samples for suspended solids. Samples were pipetted 2 inches below the surface. Resultant suspended solids were then plotted against time, as shown on the graphs.

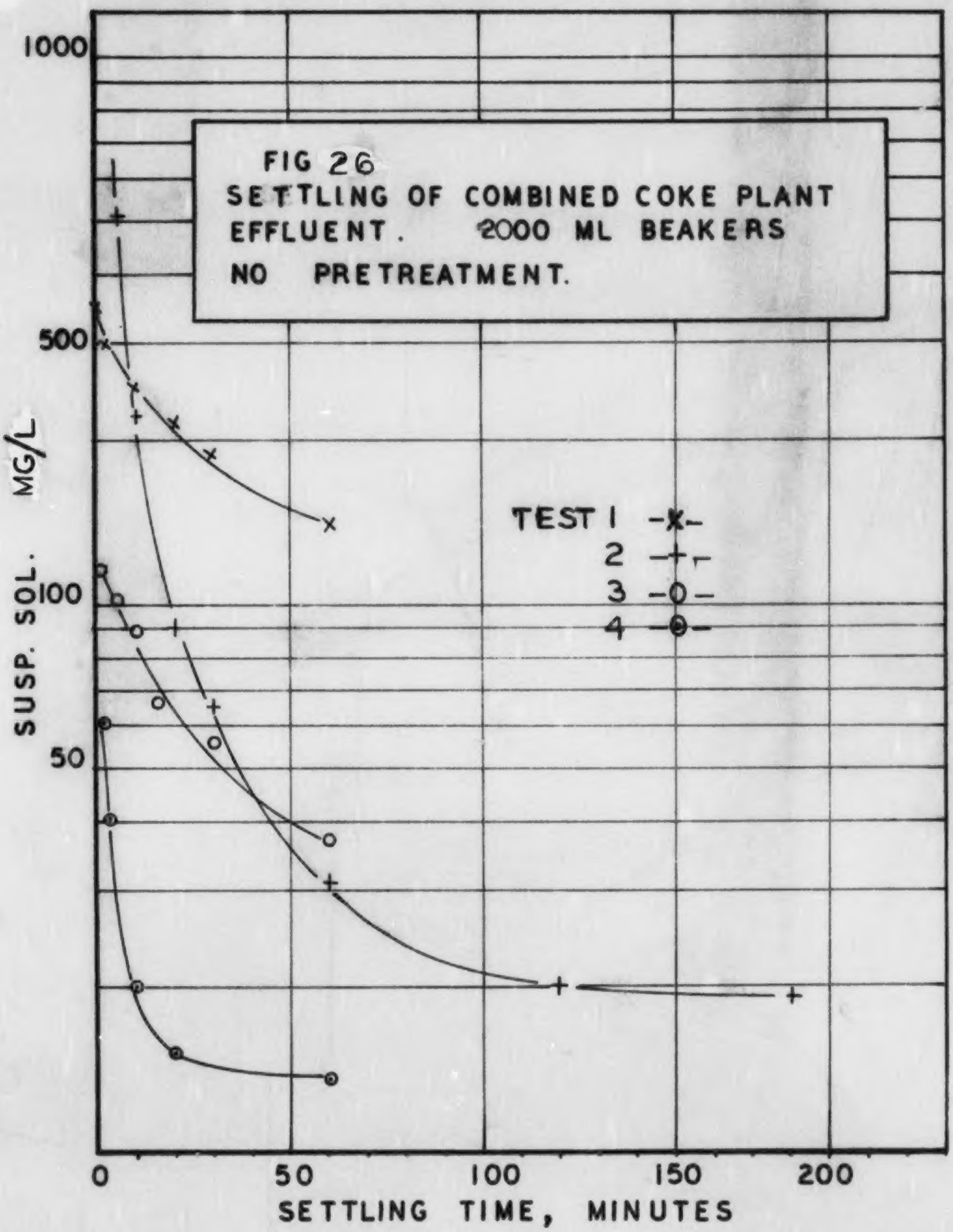


FIG 26
SETTLING OF COMBINED COKE PLANT EFFLUENT. 2000 ML BEAKERS
NO PRETREATMENT.

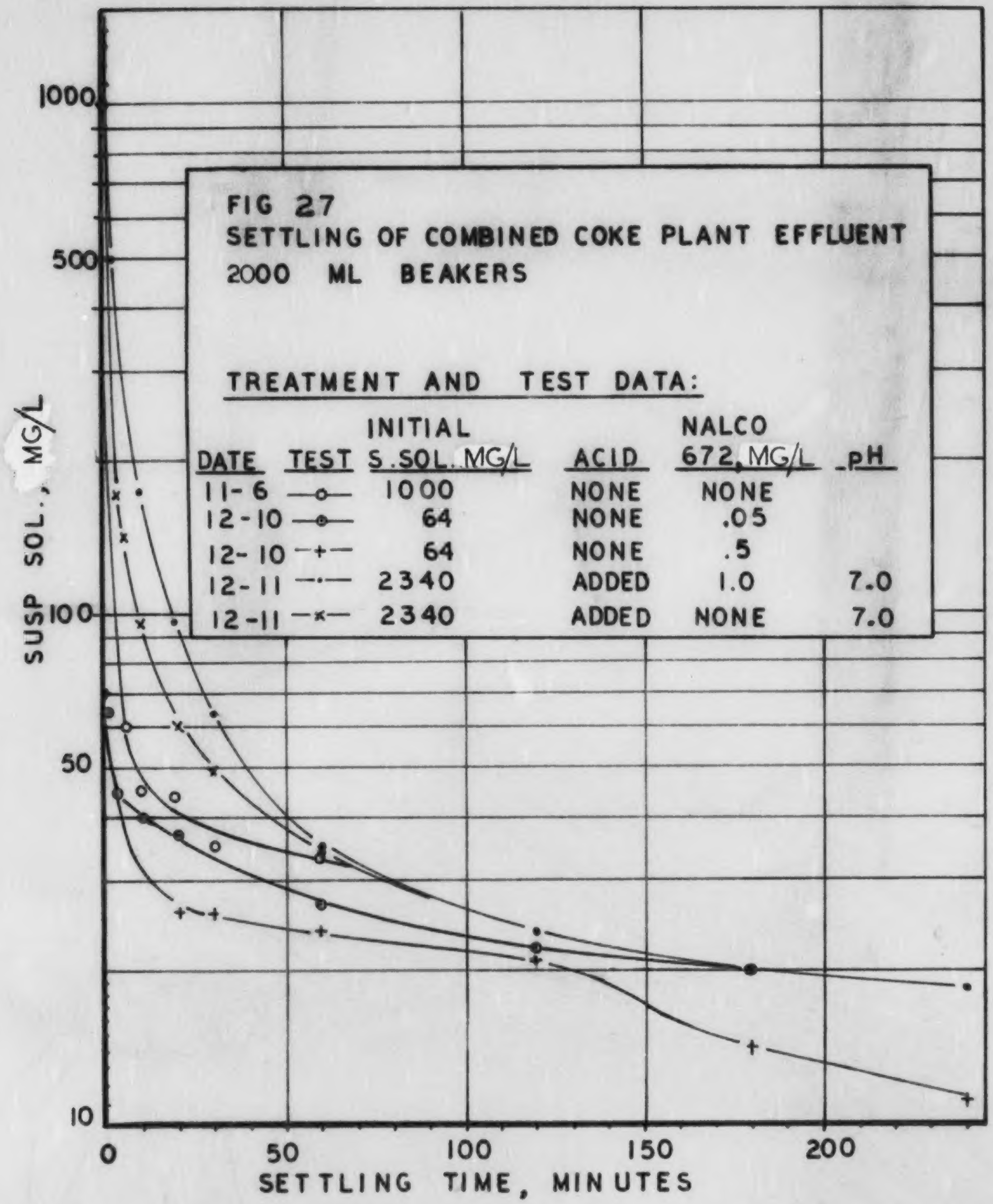


FIG 27
SETTLING OF COMBINED COKE PLANT EFFLUENT
2000 ML BEAKERS

TREATMENT AND TEST DATA:

DATE	TEST	INITIAL S.SOL. MG/L	ACID	NALCO 672, MG/L	pH
11-6	—○—	1000	NONE	NONE	
12-10	—⊙—	64	NONE	.05	
12-10	—+—	64	NONE	.5	
12-11	—·—	2340	ADDED	1.0	7.0
12-11	—×—	2340	ADDED	NONE	7.0

It is apparent that residual suspended solids levels of less than 30 mg/1 may be obtained after 1 hour of settling regardless of initial suspended solids concentrations or chemical treatment. Addition of a polyelectrolyte, Nalco 672 (nonionic), retarded rather than aided settling. Reduction in pH also seemed to hinder settling.

Two high rate pilot filters were set up at Sampling point C-7A to demonstrate removability of suspended solids present in the waste stream. A synthetic media of polyvinylchloride was used in a filter supplied by Hydromation Corporation, (a) while a filter supplied by Hayward Corporation (b) used a mixed media of synthetic material (garnet and fine sand).

A sample was pumped from Sampling point C-7A to a holding tank to provide sufficient waste water for the tests. No attempt was made to remove oil and grease. The tank was agitated to keep the solids in suspension throughout the test runs. Flows were measured by timing the flow into a 5 gallon bucket. Grab samples of the filter influent and effluent were taken for TSS at intervals of 15 and 30 minutes depending on the flow rates used during the test.

Flow rates were varied from 10-25 gpm/sq. ft. on the Hydromation filter. The results are shown in Figures 28 and 29. Figure 29 indicates that inlet total suspended solids above 100 mg/1 resulted in a filter cycle of 1 hour. The effluent solids concentrations normally varied with the influent; as the influent solids increased, an increase in effluent solids was noted. At reduced inlet TSS concentrations and flow rate, the filter merely polished the waste stream and exhibited low percent solids removal. At 20 gpm/sq. ft. the Hayward filter was unable to operate longer than 20 minutes at TSS concentrations of 70-140 mg/1 without losing flow. The effluent quality was less than 5 mg/1 TSS.

The results of the clarification test indicate that a pre-settling basin, providing 1 hour retention time and facilities for oil skimming, would provide sufficient time for acceptable solids reductions down to 30 mg/1.

(a) 39201 Amrhein Rd., Livonia, Michigan, 48150
(b) 1673 Placentia Ave., Costa Mesa, California

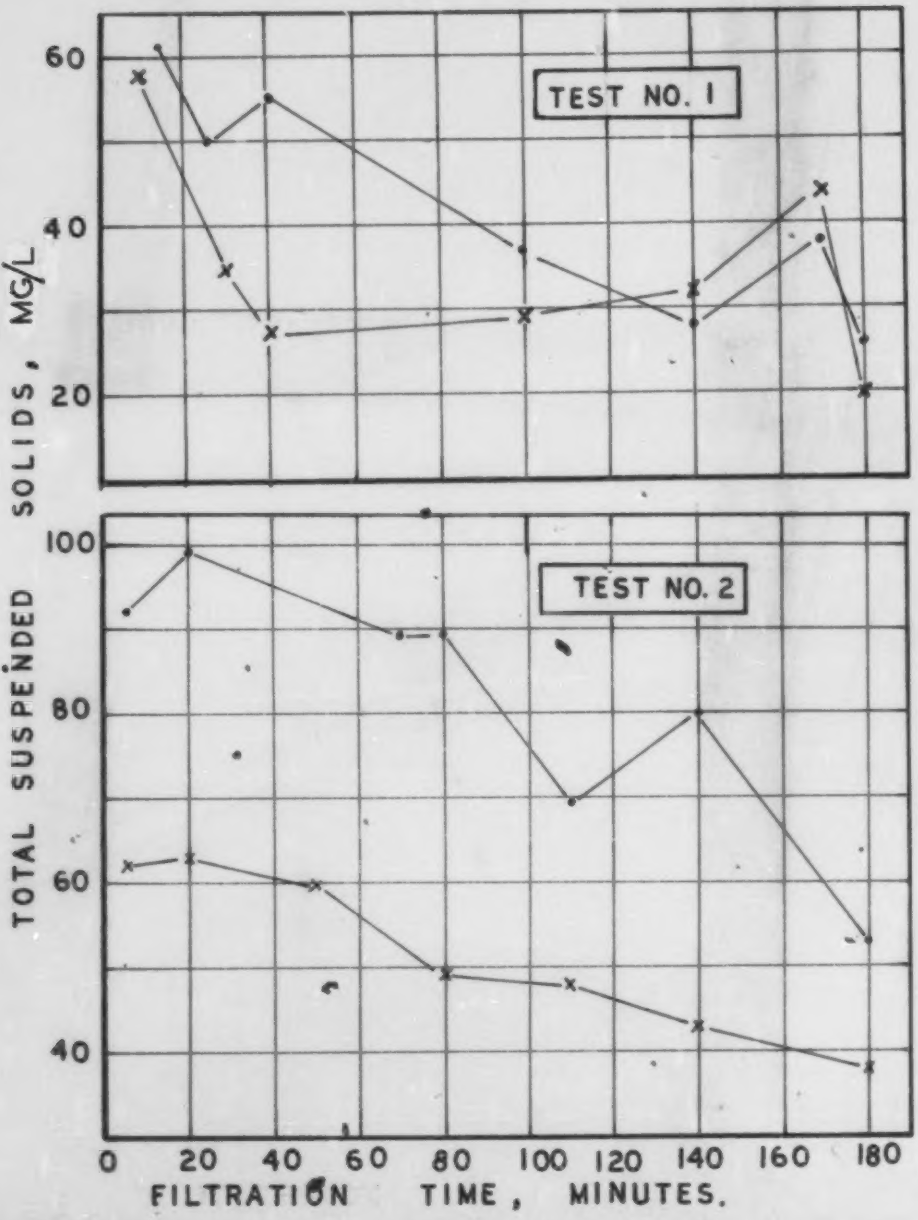


FIG. 28

RESULTS OF FILTER STUDIES AT C-7A
HYDROMATION FILTER
• INFLUENT
x EFFLUENT

TEST NO	FLOW RATE GPM/FT^2
1	25
2	22.5

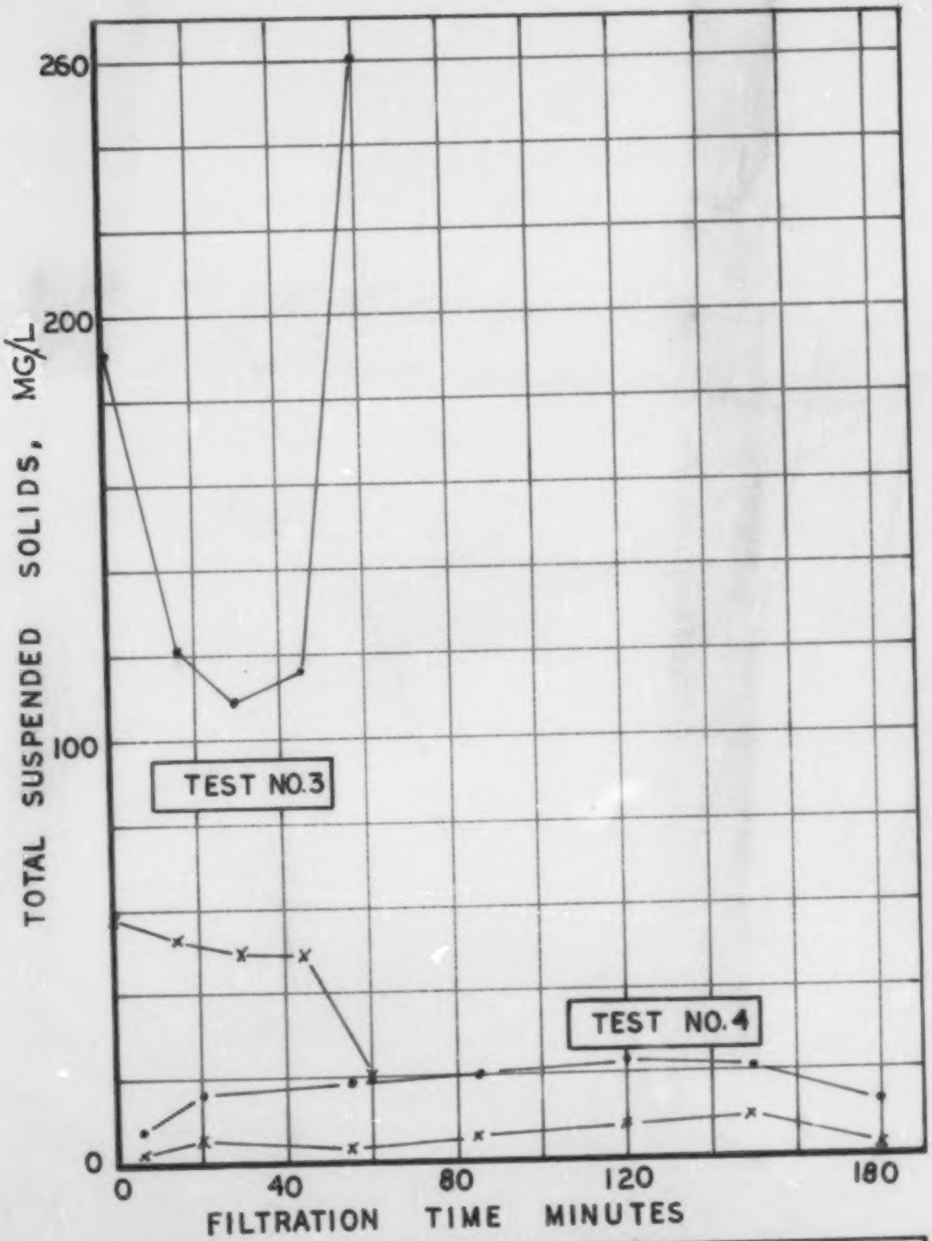


FIG 29
RESULTS OF FILTER STUDIES AT C-7A
HYDROMATION FILTER
• INFLUENT
x EFFLUENT

TEST NO	FLOW RATE GPM FT^2
3	15
4	15

LIME TREATMENT, STEAM AND AIR STRIPPING OF AMMONIA

The effect of lime treatment of ammonia flushing liquor was investigated in the laboratory and in a pilot plant steam stripping column operated at various hydraulic loadings. The pilot tests were, in essence, a continuation of the laboratory studies reported earlier. The samples were obtained from the same source and the methods of treatment were essentially the same. The system was assembled and operated in accordance with the instructions shown in Appendix A & B, pages 244-251.

Purpose

1. To investigate the removal of ammonia, cyanide, phenol, and COD by steam stripping.
2. To investigate the effects of pH on the removal of ammonia and cyanide in the steam stripping column.
3. To investigate the operating and control parameters of steam stripping.
4. To investigate the possibility of color removal by lime addition.

Procedures

See Figures 30-31 for pilot plant layout.

The pilot plant was operated by filling the feed tank with the liquor to be stripped, pumping the test solution through the column in the downward direction, and allowing the stripping steam to flow upward, counter-current to the liquid flow, at various flow rates. Rates of water and steam flow were limited by the flooding point of the column. Several rates of water and steam were used in order to determine the condition required for maximum gas removal.

The system was allowed to operate for a few minutes until equilibrium had been established before any samples were taken. Back pressures at the bottom of the column, where the effluent samples were collected, were maintained by a liquid leg on the effluent line. As shown in Figure 31 this can be accomplished by proper selection of the valve in the discharge valve assembly.

Observation of steam pressures, water flow rates, top and bottom column temperatures, and pressure at the bottom of the column were made

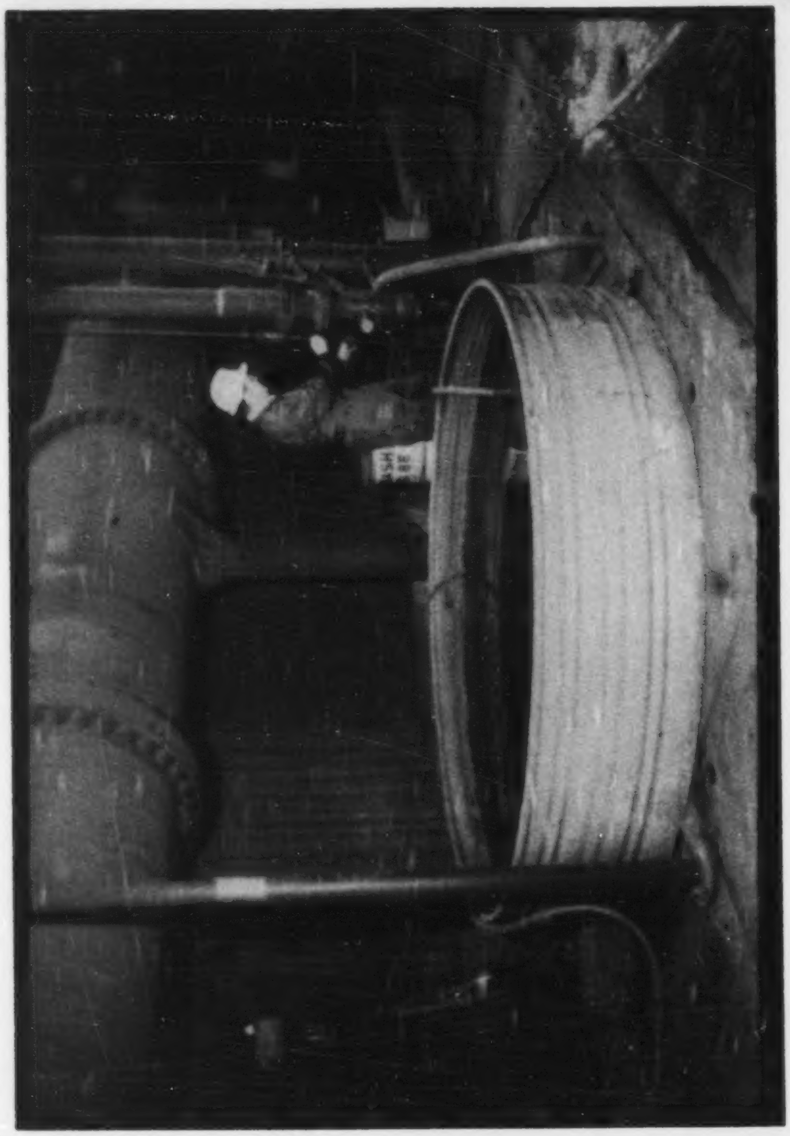


Figure 30. PILOT STEAM-AIR STRIPPING COLUMN

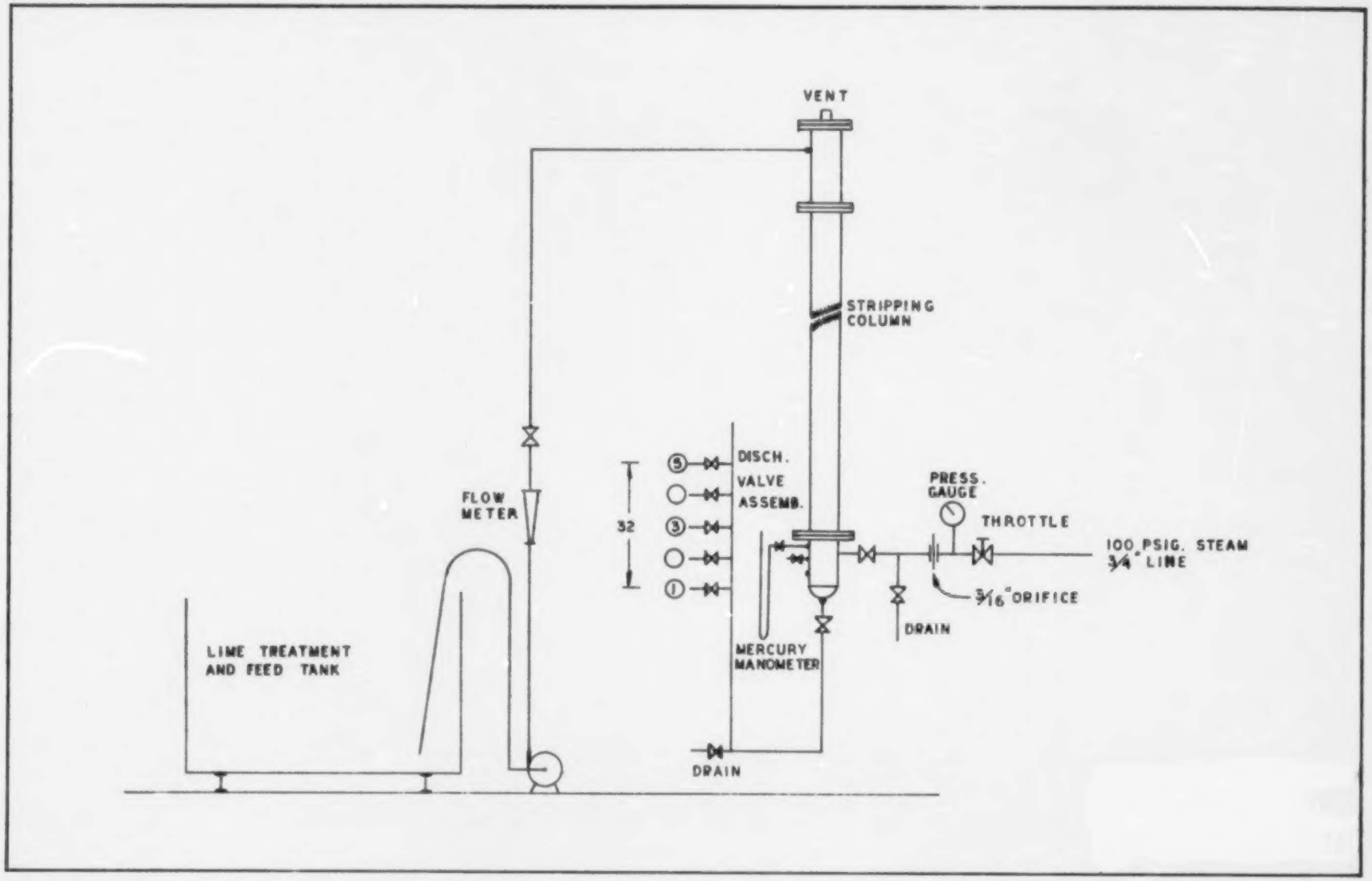

VENT
STRIPPING COLUMN
DISCH. VALVE ASSEMB.
⑤
③
①
32
FLOW METER
PRESS. GAUGE
THROTTLE
100 PSIG. STEAM 3/4" LINE
3/16" ORIFICE
DRAIN
MERCURY MANOMETER
DRAIN
LIME TREATMENT AND FEED TANK

Figure 31. STRIPPING COLUMN PILOT PLANT LAYOUT

in accordance with the operating instructions (Appendix B, page 247.)

Lime treatment was carried out in the feed tank by addition of dry plant lime and manual agitation of the tank. Quantity of lime used for each treatment was determined by the pH of the solution. Lime additions were made to obtain the desired pH. No polyelectrolyte was used in the treatment.

After addition of lime, the solution was allowed to settle for approximately 45-60 minutes. The pump suction consisted of a pipe assembly running over the top of the tank to an adjustable level in the tank. Suction was always taken from a liquid level above the level of the sludge generated by the lime treatment.

Analyses for ammonia, with the exception of those described in Table 46 series 2, were run in accordance with the coke plant procedures. Since the procedure calls for a kjeldahl digestion, all ammonia values reported should be considered as kjeldahl nitrogen. The second test series, described in Table 46, is an exception. These tests were run by simple distillation, with no digestion.

The ratio of organic nitrogen to ammonia nitrogen in the coke plant's ammonia flushing liquor are shown on Table 41.

Results

Table 42, shows the result of steam stripping of ammonia flushing liquor without any addition of lime for pH adjustment. Inasmuch as these were two different solutions, and since the reported ammonia is actually kjeldahl nitrogen, different levels of ammonia as obtained at high levels of steam application would be expected. As shown in Table 42, (tests 7 and 8), the lowest values of kjeldahl nitrogen obtained appeared to be somewhere around 1150-1200 mg/l. On the September 15 tests, the low level kjeldahl nitrogen was found to be around 300 mg/l. It should be noted that these two solutions had different pH levels, which may also be responsible for the difference in nitrogen remaining after stripping.

It appears necessary to apply as much as .18 lbs. of steam per pound of water in order to strip the solution to less than 400 mg/l kjeldahl nitrogen. It is also apparent that the main parameter of control of the operation of the column is the temperature at the top (vent gas). The most effective range is equal to or greater than 94°C. These observations are also supported by the data contained in the remaining tables.

TABLE 41

Ratio of Organic Nitrogen to Ammonia Nitrogen

Ammonia Flushing Liquor

Kjeldahl Total Nitrogen mg/l	Distillation Ammonia mg/l	Organic Nitrogen mg/l
1487	1440	47
1528	1454	74
1581	1522	79
1574	1478	96
1645	1539	106
1675	1574	101
1418	1317	101
1832	1736	96
1816	1750	66
1791	1715	76
1801	1720	81
1711	1645	66
1660	1599	61

TABLE 42

Results of Steam Stripping Gas Flushing Liquor

	STRIPPING OPERATIONS									EFFLUENT CHARACTERISTICS					
Test No.	Water Flow Meter Reading	Steam Press. psig	Liq. Feed °C	Tower Bottom °C	Tower Top °C	Steam Press. Tower Bottom Inches Hg.	Steam Flow lbs/min	Water Flow lbs/min	Steam use lbs/min H_2O	pH	NH_3 mg/1	CN mg/1	Phenol C_6H_5OH mg/1	COD mg/1	Color Percent Transm.
Raw Flushing Liquor September 8, 1969				-	-	-	-	-	-	8.5	2771	48	900	7250	-
1	.5	30	-	93.5	-	-	.99	7.3	.136	-	1292	5.0	743	5496	-
2	.5	45	-	94	-	-	1.3	7.3	.178	-	1156	2.5	726	5195	-
3	.5	80	-	94	-	-	2.3	7.3	.314	-	748	6.4	-	3743	-
4	1.0	30	-	94	-	-	.99	11.7	.0845	-	1955	12.8	800	6190	-
5	1.0	45	-	94.5	-	-	1.3	11.7	.111	-	1479	6.0	770	5990	-
6	1.0	60	-	95	-	-	1.7	11.7	.146	-	1292	5.0	618	5440	-
7	1.0	80	-	95.5	-	-	2.3	11.7	.196	-	1156	4.4	818	5380	-
8	1.0	95	-	96.5	-	-	2.8	11.7	.239	-	1156	3.6	622	5400	-
9	1.5	45	-	95	-	-	1.3	17.5	.074	-	2482	34.0	835	6701	-
10	1.5	60	-	95	-	-	1.7	17.5	.097	-	2414	29.6	868	6942	-
11	1.5	80	-	95.5	-	-	2.3	17.5	.131	-	2108	13.6	793	7002	-
12	2.0	30	-	73	-	-	.99	24.6	.044	-	2635	14.2	885	7150	-
13	2.0	45	-	83	-	-	1.3	24.6	.053	-	2516	14.5	810	6942	-
14	2.0	60	-	94	-	-	1.7	24.6	.069	-	2550	27.3	713	6701	-
15	2.0	80	-	94.5	-	-	2.3	24.6	.093	-	2448	24.1	810	6821	-
Raw Flushing Liquor September 15, 1969				-	-	-	-	-	-	9.25	1931	54.5	662	4825	19.8
1	.5	30	51	85	94.5	.14	.99	7.3	.136	8.35	476	6.4	484	3268	10.5
2	.5	45	51	87.5	95	.20	1.36	7.3	.186	8.0	348.5	2.5	350	2787	18.7
3	.5	80	49	79	95.5	.52	2.2	7.3	.305	7.6	272	1.9	183	1983	10.0
4	.5	95	49	78.5	95	1.00	2.6	7.3	.356	6.0	260.5	1.9	122	1661	4.6
5	1.0	30	48	78.5	63.5	.30	.99	11.7	.0845	9.0	1581	39.5	485	3752	8.3
6	1.0	45	47	91.5	94.5	.40	1.36	11.7	.116	8.75	986	19.66	422	3698	10.4
7	1.0	60	46	80.5	94.5	.70	1.74	11.7	.149	8.6	978.5	18.6	488	3670	9.2
8	1.0	80	45	83.5	94.5	1.68	2.2	11.7	.188	7.95	399	3.0	450	3484	15.9
9	1.0	95	45	83.5	-	2.40	2.6	11.7	.222	7.35	246.5	1.40	435	3748	11.0
10	1.5	45	45	50	59	.52	1.36	17.5	.078	9.15	1812	39	543	4512	25.8
11	1.5	60	45	57.5	52.5	.60	1.74	17.5	.1	9.1	1757	38	568	4716	26.6
12	1.5	80	45	78.5	44.5	.98	2.2	17.5	.125	9.05	1727	40	553	3912	26.9
13	2.0	30	44	40	40	.54	.99	24.6	.04	9.05	1839	38.5	562	4234	26.7
14	2.0	45	44	48.5	37.5	.52	1.36	24.6	.055	9.05	1798	44.2	523	4606	20
15	2.0	60	43	53.5	32.5	.52	1.74	24.6	.071	9.95	1761	38.5	510	4080	14.5
16	2.0	80	41	75.5	31.5	.94	2.2	24.6	.09	8.40	1734	37	501	3917	25.5
Flushing Liq., End of Test			41							9.05	1927			4427	

Table 43 shows results of stripping at elevated pH values (11.3 and 11.5). It is apparent, again that a given quantity of steam, 0.16 lbs. per pound of water, and a vent temperature of greater than or equal to 94°C will be required for removal of free ammonia.

Tables 44, and 45 show the results of stripping different solutions following adjustment of pH to the initial values indicated. Plant lime was used for pH correction. The main purpose of these tests was to determine removal of ammonia and cyanides as a function of pH.

Table 46 shows the results of steam stripping of two different solutions at various pH levels. In this case, the solution used for one series was treated with lime and allowed to resettle after completion of each test at a given pH level. In other words, the pH of the test solution was increased stepwise, as shown in the tables. Two stripping tests were run at each pH level.

All tests shown in Table 46 were run at vent temperatures greater than 94°C and 0.186 lbs. of steam per pound of water or higher.

As indicated earlier, the ammonia reported in Table 46, Series 2, was determined by distillation alone. These are not kjeldahl nitrogen concentrations as reported in the other tables.

Figure 32 shows the residual kjeldahl at various pH levels as a function of steam applied for stripping. Higher pH levels result in lower total nitrogen contents at steam usages of 0.16-0.20 lbs. of steam per pound of water.

Figure 33 shows data obtained from Table 46. In this figure kjeldahl nitrogen (Series 1) and total ammonia (Series 2) are plotted against steam application rate of 0.186 lbs. of steam per pound of water. As would be expected, the values shown for total ammonia are lower than the levels of kjeldahl nitrogen at pH levels of 10 or higher. However, the difference does not seem to be much in excess of 20-30 mg/l.

Figure 34 shows a plot of residual cyanide vs. pH at steam application levels in excess of 0.17 lbs. of steam per lb. of water. The values used for preparation of this plot were taken from any one of the tables.

It is apparent that cyanide removal is greatest per increment pH for the range pH 9.5-10.5. Table 42 indicates that complete removal of ammonia or kjeldahl nitrogen cannot be expected at a pH of 10 or below. However, adjustment of pH with lime to the level of 9.5 will result in residual ammonia levels below 100 ppm.

TABLE 43

Results of Lime Treatment and Steam Stripping:

pH 11.3

STRIPPING OPERATIONS

Actual Water Flow	Steam Press. psig	Temp. °C Tower Bottom	Temp. °C Tower Top	Steam Flow lbs/min	Water Flow lbs/min	Steam Use lbs/lb H_2O	Effluent Characteristics NH_3 mg/l	Effluent Characteristics CN mg/l	Effluent Characteristics Phenol C_6H_5OH mg/l	Effluent Characteristics COD mg/l
Raw Flushing Liquor							2678	48	900	6020
.79	30	94	94	.99	7.3	.136	136	38.5	768	5450
.725	45	93.5	95.5	1.3	7.3	.178	34	22.5	710	5410
.73	60	94.5	95.5	1.7	7.3	.233	trace	20.0	852	5210
.74	80	92.0	96	2.3	7.3	.315	34	20.5	782	5200
1.4	30	87	52	.99	11.7	.085	1989	25	-	5950
1.45	45	94.5	53	1.3	11.7	.111	1530	21	-	5610
1.39	60	94.5	93.5	1.7	11.7	.145	·125	22.8	899	5520
1.43	80	95.0	94.5	2.3	11.7	.197	44	23	868	5520
-	90	95.5	95	2.6	11.7	.222	37.4	22.3	855	5386
2.09	80	95	92.5	2.3	17.5	.135	501.5	22.5	827	5435
-	95	95.5	94	2.8	17.5	.16	45.9	23	852	5526

Remarks: Raw Flushing Liquor pH 8.4
Treatment tank, 7 ft. dia., 20" liq. depth; i.e., 420 gals.
Used 26.75 lbs. of plant lime to raise pH to 11.3

Results of Lime Treatment and Steam Stripping:

pH 11.5

Raw Flushing Liquor		-	-	-	-	-	647	-	-	-
Lime treated Liquor		-	-	-	-	-	661	14.8	-	-
.5	30	82.5	95.5	.99	7.3	.136	57	-	-	-
.5	45	82.5	95.5	1.3	7.3	.178	42	-	-	-
.5	60	83	95.5	1.7	7.3	.233	44	5.0	-	-
1.0	45	84.5	95	1.3	11.7	.111	165	-	-	-
1.0	60	86.5	95	1.7	11.7	.145	53	-	-	-
1.0	90	86	95.5	2.6	11.7	.222	28	5.27	-	-
1.5	60	81.5	94.5	1.7	17.5	.097	207	-	-	-
1.5	80	86.5	94.5	2.3	17.5	.135	77	-	-	
1.5	95	86	95	2.8	17.5	.160	75	6.13	-	-

Remarks: Raw Flushing Liquor Characteristics: pH 8.55
Raw Liquor Treated with Lime to pH 11.5

TABLE 44

Results of Lime Treatment and Steam Stripping: pH 10.0

STRIPPING OPERATIONS

Water Flow Meter Reading	Steam Press. psig	Temp., °C Tower		Steam Flow lbs/min	Water Flow lbs/min	Steam Use lbs/lb H_2O	Effluent Characteristics		
		Bottom	Top				NH_3 mg/l	CN mg/l	pH
Raw Flushing Liquor		-	-	-	-	-	-	48.5	8.3
Lime Treated Liquor		-	-	-	-	-	2720	43.9	10.0
.5	30	88	88	.99	7.3	.136	204	16.3	9.8
.5	45	91	95	1.36	7.3	.186	136	13.1	9.8
.5	60	91	95	1.74	7.3	.238	17	14.1	10.1
1.0	45	91	91	1.36	11.7	.116	478	36	9.55
1.0	60	91	93	1.74	11.7	.149	152	35.7	9.4
1.0	90	93	94	2.52	11.7	.215	17	26.3	9.45
1.5	60	80	74	1.74	17.5	.1	2074	31.6	9.8
1.5	80	83	91	2.2	17.5	.125	301	33.1	9.55
1.5	95	78	91	2.6	17.5	.149	68	30.5	9.6

Results of Lime Treatment and Steam Stripping: pH 10.5

Raw Flushing Liquor		-	-	-	-	-	2380	46.5	8.5
Lime Treated Liquor		-	-	-	-	-	2227	44.5	10.5
.5	30	91	95	.99	7.3	.136	136	36	10.2
.5	45	91	95	1.36	7.3	.186	64	36	10.5
.5	60	91	95	1.74	7.3	.238	64	31	10.5
1.0	45	91	89	1.36	11.7	.116	1782	41	10.5
1.0	60	90	93	1.74	11.7	.149	103	35	10.5
1.0	90	-	95	2.52	11.7	.215	64	35.3	10.5
1.5	60	78	85	1.74	17.5	.1	1564	38.2	10.5
1.5	80	84	91	2.2	17.5	.125	273	39.9	10.5
1.5	95	-	-	2.6	17.5	.149	64	33.4	10.25

TABLE 45

Results of Lime Treatment and Steam Stripping:

pH 11.0

STRIPPING OPERATIONS

Water Flow Meter Reading	Steam Press. psig	Temp., °C Tower Bottom	Top	Steam Flow lbs/min	Water Flow lbs/min	Steam Use lbs/lb H_2O	Effluent Characteristics NH_3 mg/l	CN mg/l	pH
Raw Flushing Liquor		-	-	-	-	-	2480	47.4	8.6
Lime Treated Liquor		-	-		-	-	2380	46.2	11.0
.5	30	93	95	.99	7.3	.136	17	30.5	-
.5	45	93	95	1.36	7.3	.186	17	34.5	-
.5	60	93	95	1.74	7.3	.238	trace	34	-
1.0	45	91	92	1.36	11.7	.116	612	36.8	-
1.0	60	92	93	1.74	11.7	.149	102	36.8	-
1.0	90	-	95	2.52	11.7	.215	trace	36.5	-
1.5	60	80	65	1.74	17.5	.1	1666	38.2	10.9
1.5	80	84	92	2.2	17.5	.125	34	36	-
1.5	95	86	93	2.6	17.5	.149	-	33	-

Results of Lime Treatment and Steam Stripping:

pH 11.5

Raw Flushing Liquor		-	-	-	-	-	2584	45.2	8.4
Lime Treated Liquor		-	-	-	-	-	2380	38.8	11.5
.5	30	91	95	.99	7.3	.136	34	29	11.4
.5	45	91	95	1.36	7.3	.186	34	30.5	-
.5	60	91	95	1.74	7.3	.238	34	27.5	-
1.0	45	90	91	1.36	11.7	.116	612	29.2	
1.0	60	91	94	1.74	11.7	.149	34	26.5	
1.0	90	85	95	2.52	11.7	.215	34	29.5	11.4
1.5	60	80	52	1.74	17.5	.1	1802	31.5	11.5
1.5	80	83	92	2.2	17.5	.125	34	30.5	11.4
1.5	95	85	93	2.6	17.5	.149	66	30.7	

TABLE 46

Results of Steam Stripping Gas Flushing Liquor

Series 1

STRIPPING OPERATIONS

Water Flow Meter Reading	pH Feed	Steam Press. psig	Temp., °C Tower Bottom	Temp., °C Tower Top	Steam Flow lbs/min	Water Flow lbs/min	Steam Use lbs/lb H_2O	Effluent Characteristics NH_3 mg/l	Effluent Characteristics CN mg/l
Raw Flushing Liquor pH 8.7			-	-	-	-	-	1700	72
.5	8.7	45	85	95	1.3	7.3	.186	136	2.1
1.0	8.7	80	84.5	95	2.3	11.7	.197	136	2.4
.5	9.5	45	93	95	1.3	7.3	.186	68	4.6
1.0	9.5	80	85.5	95.5	2.3	11.7	.197	30	5.3
.5	10.0	45	94	95	1.3	7.3	.187	42	31.2
1.0	10.0	60	94	94.5	2.3	11.7	.197	44	28.5
.5	10.5	45	82.5	95	1.3	7.3	.186	48	38.4
1.0	10.5	60	86.5	95	2.3	11.7	.197	48	41
.5	11.3	45	82.5	95	1.3	7.3	.186	43	39
1.0	11.3	80	82.5	95.5	2.3	11.7	.197	41	41.6

Results of Steam Stripping Gas Flushing Liquor

Series 2

Water Flow Meter Reading	pH Feed	Steam Press. psig	Temp., °C Tower Bottom	Temp., °C Tower Top	Steam Flow lbs/min	Water Flow lbs/min	Steam Use lbs/lb H_2O	Effluent Characteristics NH_3 mg/l	Effluent Characteristics CN mg/l
Raw Flushing Liquor pH 8.6			-	-	-	-	-	1768	54.5
.5	8.6	45	77.5	95	1.3	7.3	.186	298	4.6
1.0	8.6	80	88	94.5	2.3	11.7	.197	259	3.0
.5	9.7	45	80	94.5	1.3	7.3	.186	91	6.1
1.0	9.7	80	81.5	94.5	2.3	11.7	.197	115	7.4
.5	10.0	45	68	94.5	1.3	7.3	.186	17	17.1
1.0	10.0	80	86.5	94.5	2.3	11.7	.197	18	15.6
.5	10.5	45	93.5	94.5	1.3	7.3	.183	12	34.6
1.0	10.5	80	92.5	94.5	2.3	11.7	.197	23	35.8
.5	11.3	45	93.5	94.5	1.3	7.3	.183	22.5	36
1.0	11.3	80	93.5	94.5	2.3	11.7	.197	22.5	36

Remarks: Total ammonia as NH_3

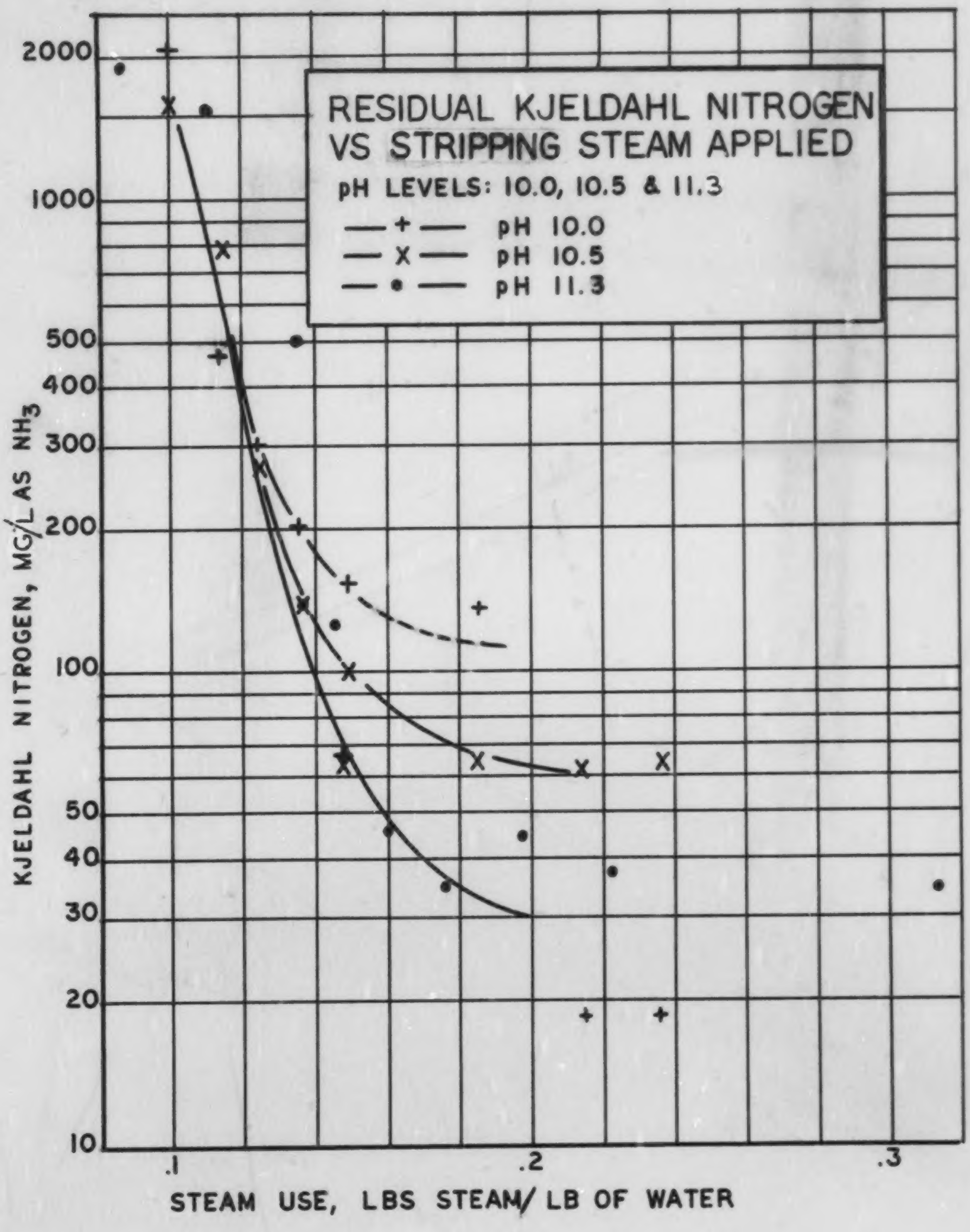

RESIDUAL KJELDAHL NITROGEN VS STRIPPING STEAM APPLIED
pH LEVELS: 10.0, 10.5 & 11.3
—+— pH 10.0
—X— pH 10.5
—●— pH 11.3
KJELDAHL NITROGEN, MG/L AS NH_3
2000
1000
500
400
300
200
100
50
40
30
20
10
.1
.2
.3
STEAM USE, LBS STEAM/ LB OF WATER

FIG. 32

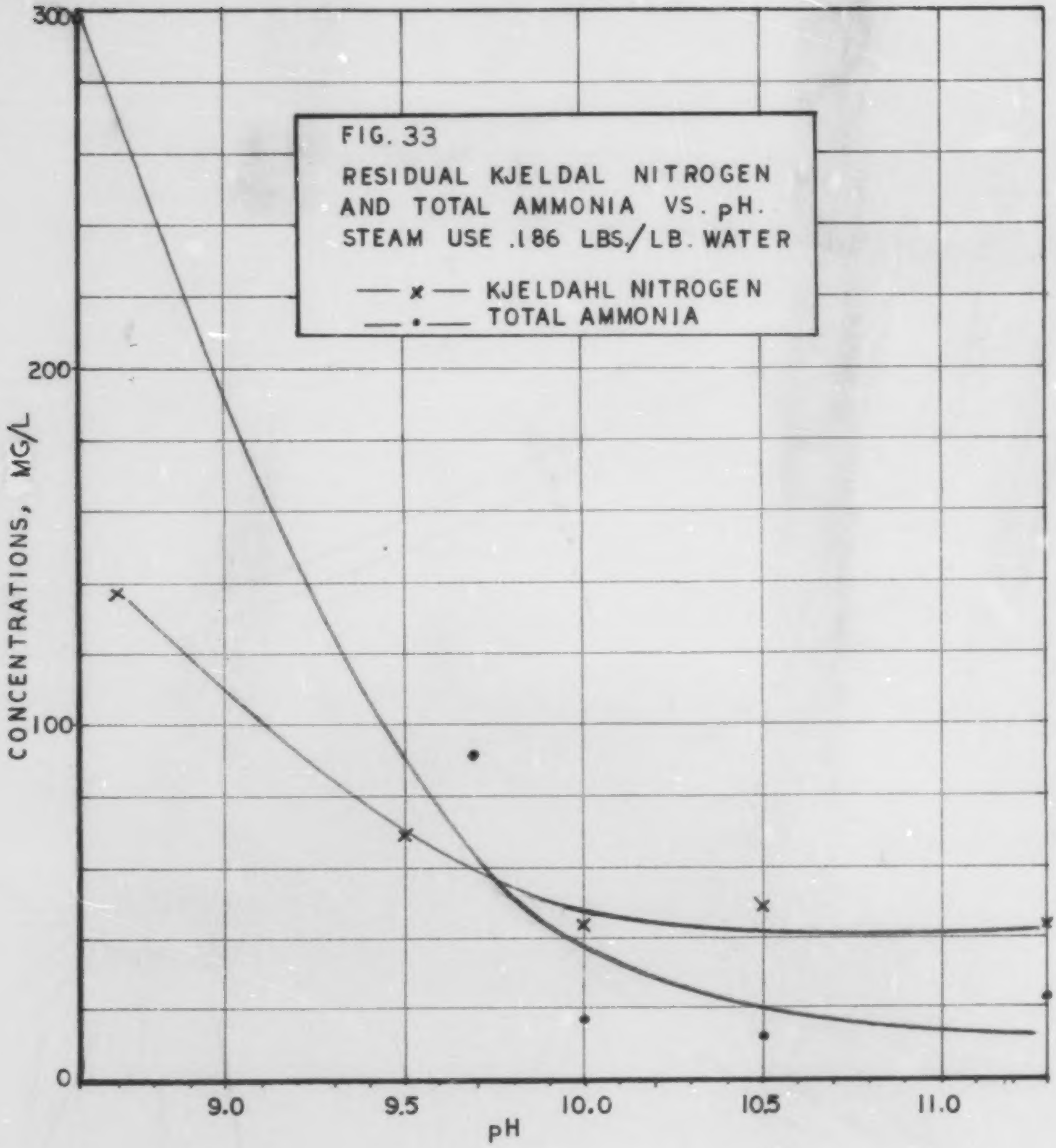
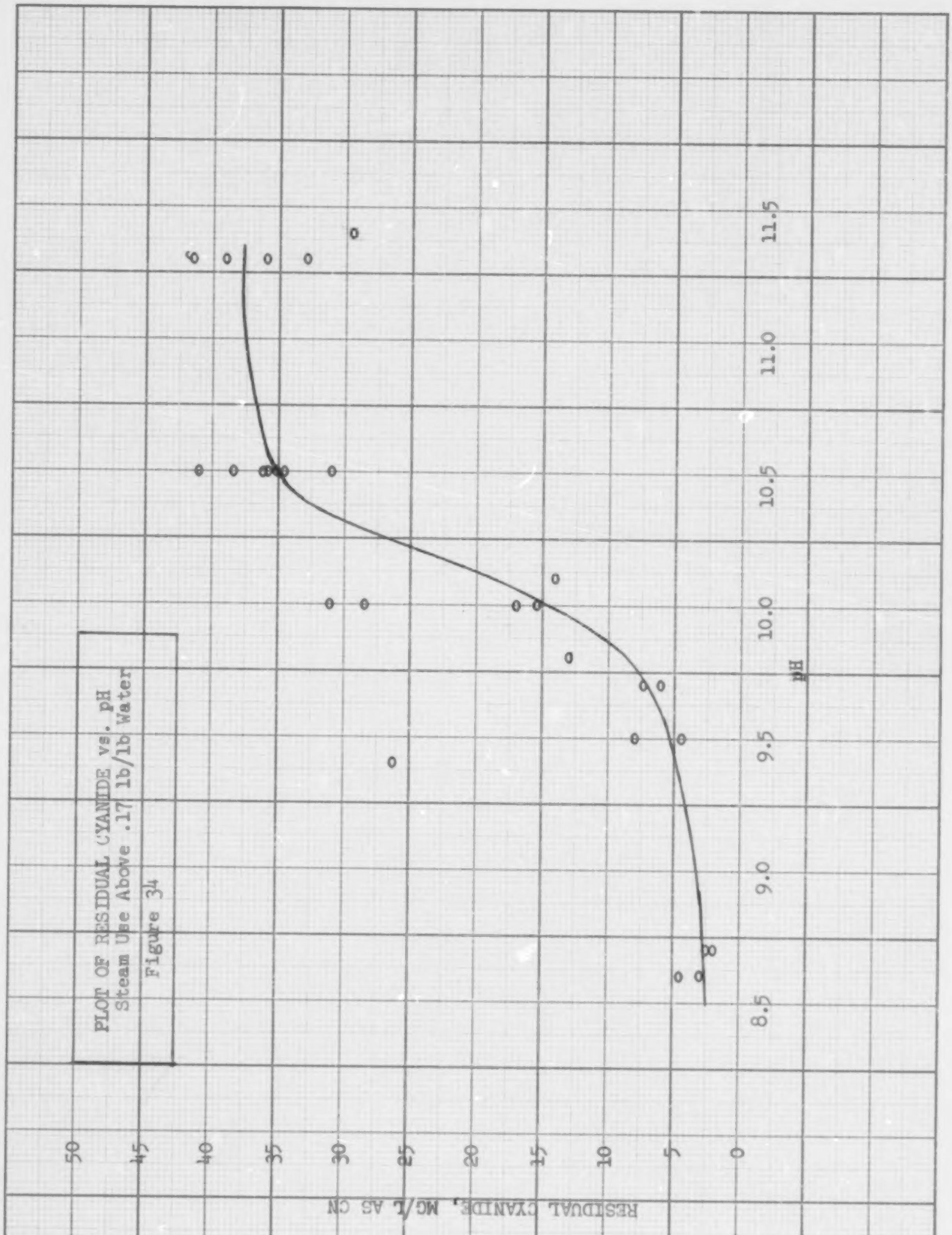


FIG. 33
RESIDUAL KJELDAL NITROGEN AND TOTAL AMMONIA VS. pH. STEAM USE .186 LBS./LB. WATER
—x— KJELDAHL NITROGEN
—•— TOTAL AMMONIA
CONCENTRATIONS, MG/L
300
200
100
0
9.0
9.5
10.0
10.5
11.0
pH



PLOT OF RESIDUAL CYANIDE vs. pH
Steam Use Above .17 lb/lb Water
Figure 34

Considering the effect of ammonia and cyanide as pollutants, it would appear from this data that adjustment of pH to approximately 9.5 would result in optimum removal of the two constituents by steam stripping. Further increase in pH would impair the removal of cyanides.

Color removal was measured by using a Hach color filter 5543, with distilled water as a blank. Samples were analyzed following steam stripping and lime addition. Results show that there was little or no color removal obtained by either steam stripping or lime addition. Table 42, page 130 contains color data for steam stripping.

Tables 47-48, show results of air stripping. Steam was used for heating in the feed tank to the top and bottom temperatures shown. Increase in pH, column temperature, and air rate result in better ammonia removal. Ultimate removal as determined in earlier tests at near boiling temperatures at the top of the tower could not be determined in this series because of liquid overflow from the vent.

Table 48, shows increasing ammonia removal by air stripping at increasing pH levels. Analyses were made using simple distillation.

The removal of cyanides accomplished during the steam stripping tests should be disregarded as a method of reducing the concentration of this pollutant from the total coke plant waste effluent. By discharging the stripped gases back into the gas stream for ammonia recovery, the cyanides would be picked up again in the cooling water used at the final coolers.

The calculated cost of steam for stripping excess ammonia flushing liquor is $71 per day using a flow of 60 gpm and a steam cost of 53¢/1000 lbs. of steam, or $0.082/1000 gallons.

Lime costs calculated for raising the pH to remove fixed ammonia are $9.27 per day or $.011/1000 gallons (based on a flow of 60 gpm, CF&I lime cost of .0065¢/lb., and 16.5 lbs. needed to raise the pH of 1000 gallons to pH 11.5). At the same flow, daily costs would be $5.05 to raise the pH to 9.5.

TABLE 47

Results of Air Stripping Gas Flushing Liquor

12-18-69 Series 1

Flow			Temperatures °F			Ammonia NH_3 mg/l	
Water GPM	Air CFM	pH	Tank	Bottom Column	Top Column	Inlet	Outlet
-	-	9.3	-	-	-	1697	-
.88	5	9.3	56	40	38	-	1638
-	5	9.3	56	40	38	-	1624
-	5	9.3	56	40	38	-	-
-	5	9.3	56	40	38	-	1642
-	15	9.3	56	40	38		1460
-	15	9.3	56	40	38		1482
-	15	9.3	56	40	38	-	1519
-	15	9.35	56	40	38		1491

Results of Air Stripping Gas Flushing Liquor

12 18-69 Series 2

Water GPM	Air CFM	pH	Tank	Bottom Column	Top Column	Inlet	Outlet
-	-	9.3	96	85	83	1642	
.88	5	9.3	96	85	83		1568
-	5	9.3	96	85	83		1568
-	5	9.3	96	85	83	-	1561
-	5	9.3	96	85	83	-	1569
-	15	9.3	96	85	83	-	1220
-	15	9.3	96	85	83	-	1207
-	15	9.3	96	85	83	-	1196
-	15	9.3	96	85	83	-	1196

TABLE 48

Results of Air Stripping Gas Flushing Liquor

12-18-69
Series 1

Flow			Temperatures °F			Ammonia Nitrogen mg/l		Cyanide mg/l	
Water GPM	Air CFM	pH	Tank	Bottom Column	Top Column	Inlet	Outlet	Inlet	Outlet
-	-	8.8	108	-	-	1511	-	-	-
.88	15	8.9	108	65	98	-	1393	-	-
-	15	8.9	108	65	98	-	1393	-	-
-	25	8.8	118	66	112	-	1262	-	-
-	25	8.9	118	66	112	-	1299	-	-
-	35	8.95	118	65	108	-	1171	-	-
-	35	9.05	118	65	108	-	1161	-	-
-	45	-	113	60	84	-	1114	-	-
-	45	-	113	60	84	-	1097	-	-

Results of Air Stripping of Flushing Liquor
Series 2

	Water GPM	Air CFM	pH	Tank	Bottom Column	Top Column	Ammonia Inlet	Ammonia Outlet	Cyanide Inlet	Cyanide Outlet
12-22-69	-	-	-	86	-	-	1568	-	92.6	-
	.88	40	-	86	56	76	-	1359	-	38.4
	-	40	-	86	56	76	-	1430	-	54.4
	-	40	-	95	60	80	-	1268	-	20.4
	-	40	-	95	60	80	-	1228	-	16.3
12-23-69	-	-	9.2	68	-	-	1255	-	42	-
	.88	30	9.4	68	40	55	-	1128	-	24
	-	30	9.3	68	40	55	-	1111	-	24
	-	-	9.25	77	-	-	1294	-	50	-
	.88	30	9.15	77	48	60	-	1147	-	20
	-	30	9.2	77	48	60	-	1127	-	21
(pH 9.5)	-	-	9.5	68	-	-	1514	-	-	-
	.88	35	9.5	68	50	40	-	1275	-	-
	-	35	9.5	68	50	40	-	1054	-	-
(pH 10.0)	-	-	9.6	68	-		1514		-	-
	.88	35	9.6	68	52	42	-	584	-	
	-	35	9.6	68	52	42	-	605	-	
(pH 11.0)	-	-	11.2	86	-	-	1470	-	75	-
	.88	35	11.2	86	50	52	-	861	-	50
	-	35	11.2	86	50	52	-	750	-	50
(pH 12.0)	-	-	11.8	68	-	-	1514	-	60	-
	.88	35	11.8	68	55	42		731	-	50
	-	35	11.8	68	55	42	-	675	-	49

Remarks: Steam used for heating in feed tank.

ION EXCHANGE

Use of ion exchange techniques for ammonia removal has been investigated by a number of researchers concerned with the problem of ammonia removal from secondary sewage treatment plant effluents (3). It was felt that a process involving ion exchange might be used at the coke plant since the spent acid from the regeneration of the ion exchange units could be used in the ammonia saturator.

A study was made of the performance of a strongly acidic and a weakly acidic cation exchanger in order to determine the degree of removal and the feasibility of incorporating such a process in the final treatment system.

The actual testing was done in Chicago in Nalco's Ion Exchange Laboratory. A 55 gallon drum containing ammonia flushing liquor was shipped to Chicago for use in this study.

The purpose of this study was to determine the ammonia removal from flushing liquor with ion exchange resins such as the strongly acidic cation resin, Amberlite 200, H^+ form, and a weakly acidic resin such as Amberlite IRC-84, H^+ form.

Resin data can be found in Appendix K, page 287.

Column Operation of Ion Exchange Resins Amberlite 200 and Amberlite IRC-84

The flushing liquor was passed through the cation exchange resins Amberlite 200 and Amberlite IRC-84 H^+ forms, at a flowrate of 2 gpm/ft^3. The run was discontinued when the effluent analysis showed a 50 mg/l increase in ammonia leakage. The effluent was also analyzed for TOC, cyanide, color, free mineral acidity, pH and COD.

Procedure

Removal of Ammonia from Flushing Liquor

Apparatus and Equipment

a. 1" ID Lucite column with Saran screen bottom

b. Eastern Industries pump

c. Beckman pH meter

d. B&L "spectronic 20" Spectrophotometer

e. Flowraters

f. N/50 standard acid

g. Special indicator Nalco No. 260

Preparation of Feed Solution

The flushing liquor received in a 55 gallon steel drum was strained through glass wool to remove particles of dirt or coal. These particles were removed to avoid column plugging.

Analysis of Flushing Liquor from CF&I

a.	Ammonia as NH_3	1400 mg/l
b.	Cyanide as CN	48 mg/l
c.	Phenols as Phenol	640 mg/l
d.	Total Organic Carbon (TOC)	1600 mg/l
e.	Chemical Oxygen Demand (COD)	4500 mg/l
f.	Total Hardness as $CaCO_3$	27 mg/l
g.	Sodium as Na_2O	11 mg/l
h.	pH	9.5

Column Operation of Ion Exchange Resins

The flushing liquor was pumped through the 1" ID Lucite column at a flowrate of 2 gpm/ft^3. One bed volume was equal to 4.7 gallons. Effluent samples were collected at regular intervals for NH_3, CN, phenols, TOC, pH and COD analyses. The cycle was discontinued when the ammonia in the effluent showed an increase of 50 mg/l or more and was regenerated prior to the start of the 2nd cycle.

Preparation of Amberlite 200 and IRC-84

The Amberlite 200 and Amberlite IRC-84 were regenerated with 6 lbs./cu. ft. of H_2SO_4 and demineralized water rinsed prior to the exhaustion run. The Amberlite IRC-84 was also conditioned prior to the regeneration. The conditioning consisted of treating the bed with two equivalent volumes of 5% NaCl, rinsing with demineralized water, and treating with organic leachables. Each exhaustion cycle was followed by demineralized water backwash and 6 lbs./cu. ft. regeneration with 6% H_2SO_4.

Ammonia Removal from Flushing Liquor with Ion Exchange Resins

These results are compiled in Table 49 and graphed on Figures 35 and 36. The average ammonia leakage during the Amberlite 200 run was about 100 mg/l as NH_3 or 7% of the total ammonia in the flushing liquor. Amberlite IRC-84, however, showed twice as much ammonia leakage as Amberlite 200. The ammonia removal capacity of Amberlite IRC-84 was larger as evidenced by the longer bed life before ammonia breakthrough. As noted in the Appendix, page 287, the Amberlite 200 has a total exchange capacity of 1.74 meq/ml, while the Amberlite IRC-84 has a total exchange capacity of 4.13 meq/ml.

The operating capacity of Amberlite 200 to a 150 mg/l ammonia leakage was 0.7 meq/ml or about 50% of the normal operating capacity of this material at flow rate of 2 gals/cu. ft./min. The average ammonia leakage of Amberlite IRC-84 was 200-240 mg/l and its operating capacity was 1.1 meq/ml at a 16 bed volume throughput. Therefore, the Amberlite IRC-84 shows about 60% more operating capacity than Amberlite 200. However, the ammonia leakage is twice as large. The IRC-84 also exhibited swelling of approximately 20%.

TOC Removal from Flushing Liquor with Amberlite 200 and Amberlite IRC-84

The TOC removal from flushing liquor is shown in Figures 38 & 39. Both resins show a considerably larger removal of TOC at the start of the run. At about four bed volumes of the effluent, the removal was about 50% of the TOC in the flushing liquor. Then the TOC removal was gradually decreasing to 30% at the end of the run. The TOC removal also noted by the color change of the effluent and by the color change of the resin. This was especially true with the Amberlite 200. The resin grew darker at the start of the run and then gradually becambe lighter as it became exhausted, indicating a release of the adsorbed organic color.

Chemical Oxygen Demand (COD) in the Amberlite 200 and IRC-84 Effluent

The COD of the flushing liquor effluent was reduced by both resins from 38 to 40% at the start of the run. At the end of the run the COD removal was about 30%. Data appears in Table 49.

Cyanide Reduction of Flushing Liquor with Amberlite IRC-84 and Amberlite 200

Neither one of these resins showed any cyanide removal.

TABLE 49

Data Ion Exchange Tests

Bed Volumes	Amberlite IRC-84						Amberlite 200							
	NH_3 mg/l Cycles		TOC mg/l Cycles		Phenol mg/l Cycles		NH_3 mg/l Cycles		TOC mg/l Cycles		FMA mg/l $CaCO_3$ Cycles		Phenol mg/l Cycles	
	1	2	1	2	1	2	1	2	1	2	1	2	1	2
Raw Analysis	1400		1600		640		1400		1600		-		640	
1	-	-	-	-	-	-								
2	180	200	620	660	-	440	94	-	700	700	-	-	-	-
2.5	-	-	-	-	420	-	92	92	-	-	-	-	-	480
2.9	-	-	-	-	-	-	-	-	-	-	-	-	-	-
3	-	-	-	-	-	-	-	-	-	-	-	-	560	-
3.5	-	-	-	-	-	-	-	102	-	-	300	280	-	-
4	192	180	730	740	-	-	100	-	830	-	-	-	-	-
4.5	-	-	-	-	610	530	-	-	-	780	-	-	-	-
5	204	188	-	-	-	-	-	94	-	-	-	-	650	550
5.5	-	-	-	-	-	-	95	-	-	830	220	220	-	-
6	-	-	-	-	590	620	-	-	900	-	-	-	-	-
7	-	232	900	870	-	-	-	-	-	-	-	-	690	600
8	-	-	-	900	610	600	100	100	-	860	170	180	700	600
9	-	244	910	-	-	-	-	100	1070	900	-	140	680	-
9.5	-	-	-	-	-	640	-	-	-	-	-	-	-	-
10	204	228	-	-	-	-	-	-	-	-	-	-	-	-
10.5	-	-	-	-	640	-	-	-	-	-	-	-	-	-
11	-	-	940	-	-	670	-	-	-	-	-	-	-	-
11.5	-	-	-	-	-	-	-	-	-	-	-	120	-	-
12	-	252	-	-	-	-	-	-	-	-	90	-	-	-
13	-	280	-	-	-	-	-	-	-	-	-	-	-	-
15	-	-	-	-	-	-	-	-	-	-	-	-	-	-
16	-	300	-	-	-	-	-	-	-	-	-	-	-	-
18	-	360	-	-	-	-	-	-			40	-	-	-

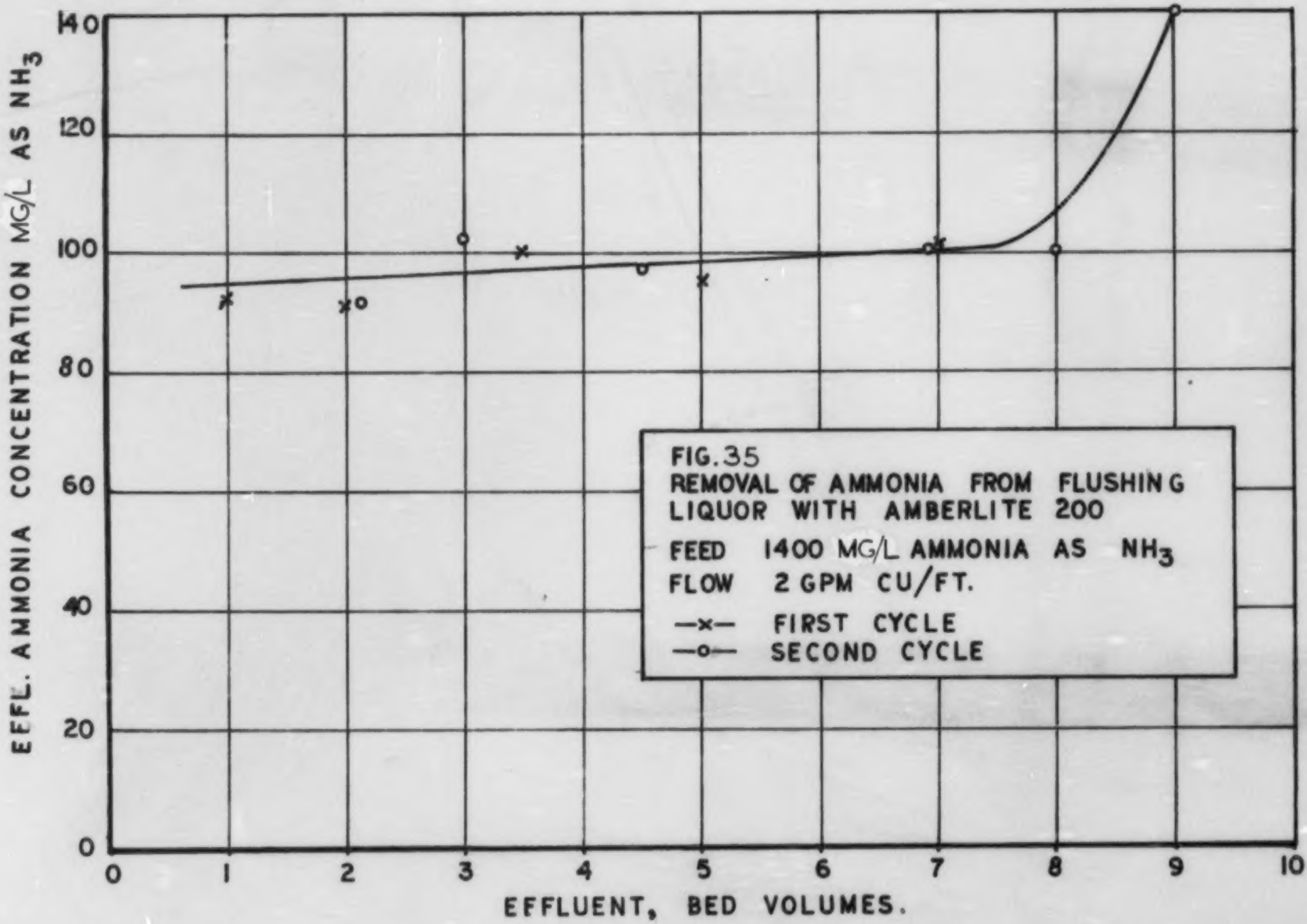
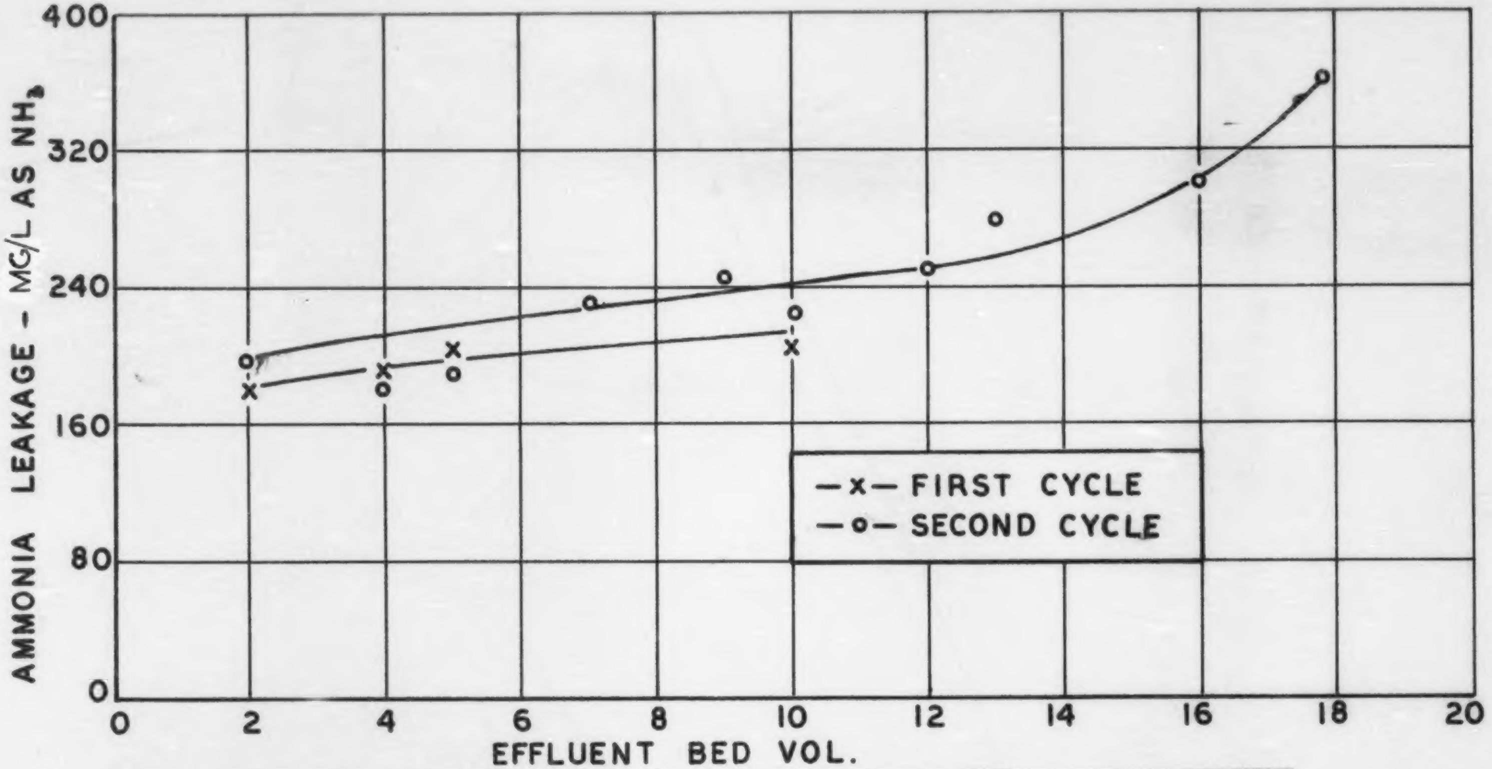


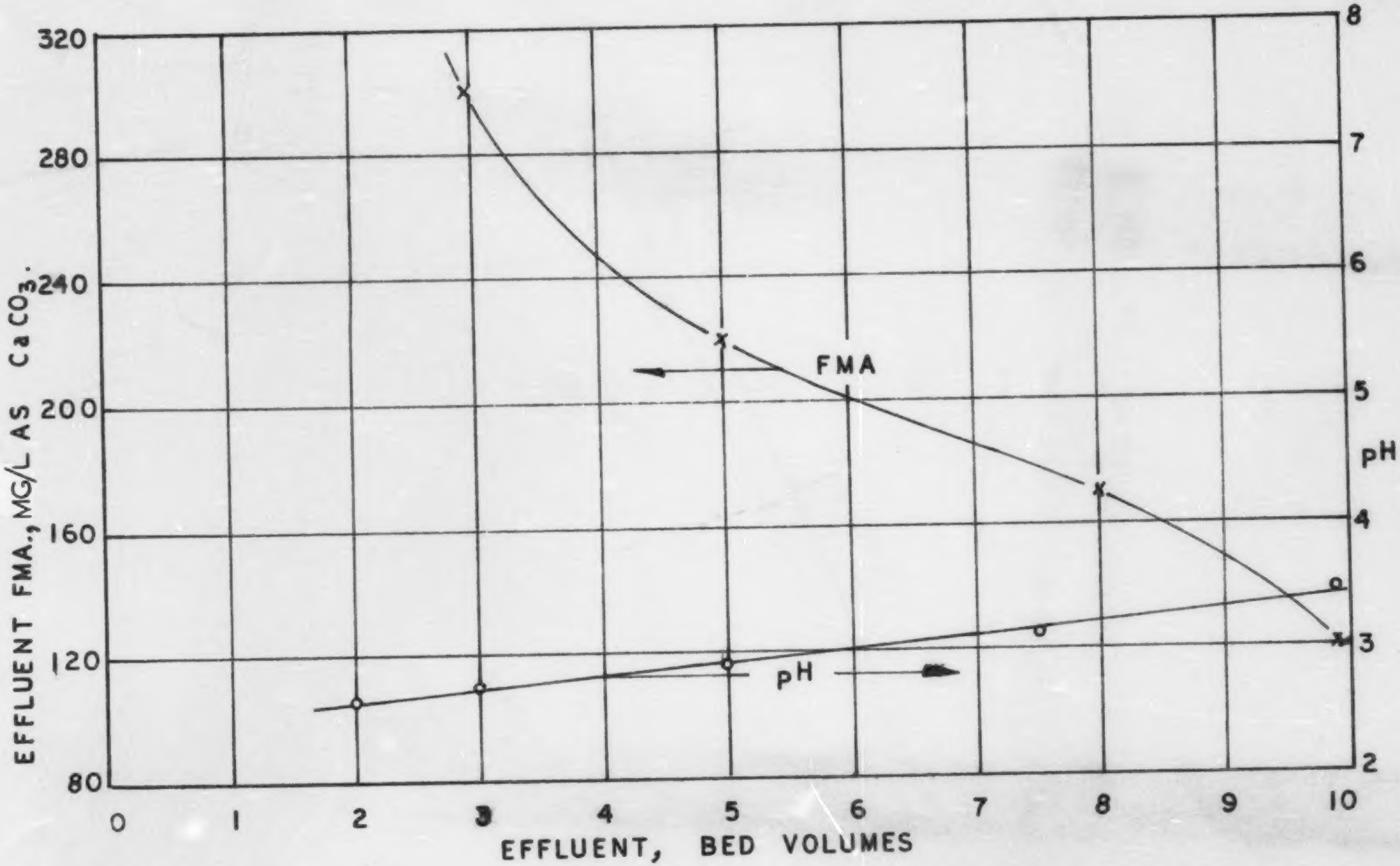
FIG. 35
REMOVAL OF AMMONIA FROM FLUSHING LIQUOR WITH AMBERLITE 200

FEED 1400 MG/L AMMONIA AS NH_3
FLOW 2 GPM CU/FT.



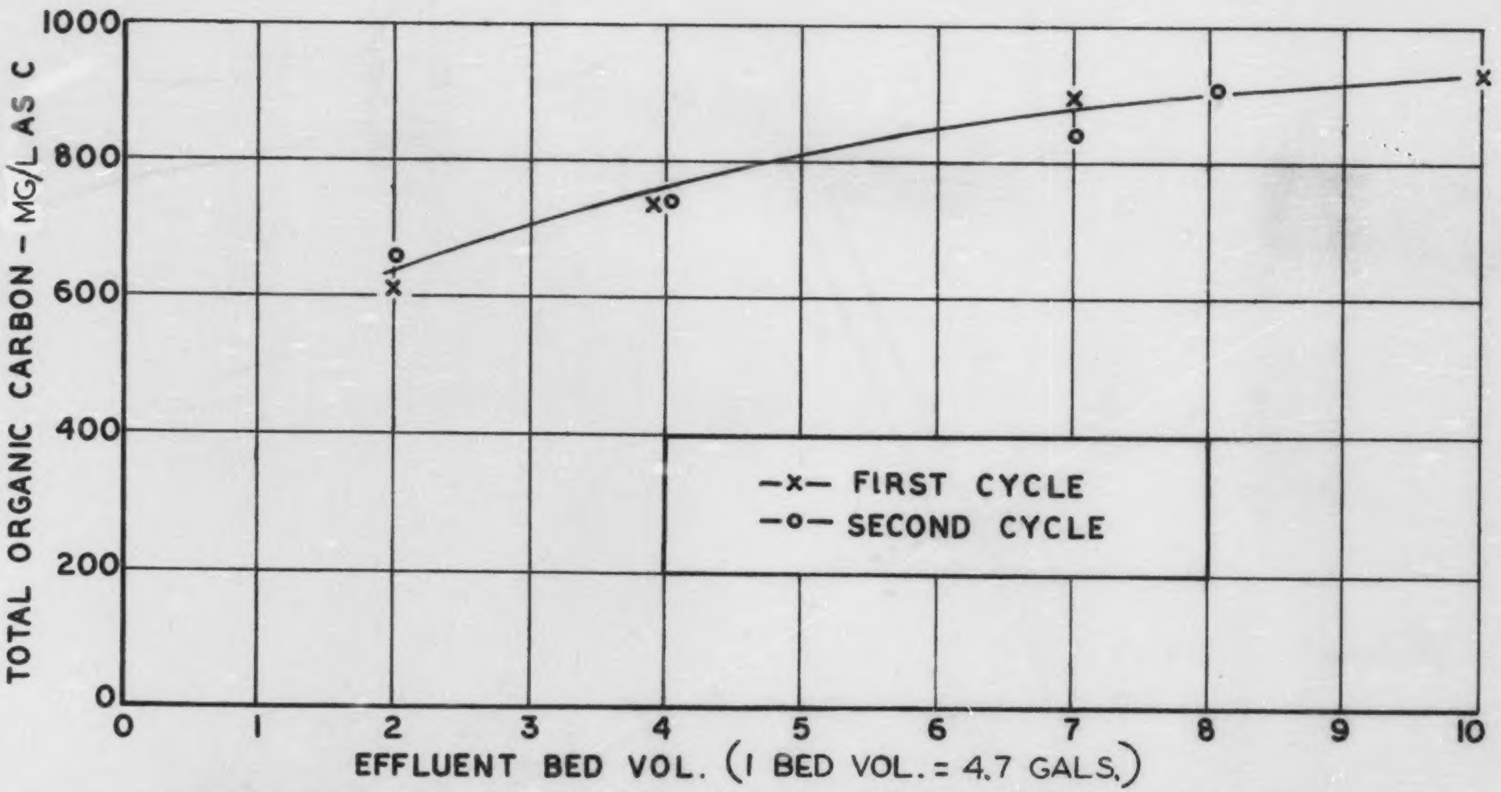
REMOVAL OF AMMONIA FROM FLUSHING LIQUOR BY AMBERLITE IRC-84

FEED: 1400 MG/L AMMONIA AS NH_3
FLOWRATE: 2 GPM / CU. FT.
FIG 36



pH AND FREE MIN. ACIDITY OF FLUSHING LIQUOR TREATED BY AMBERLITE IRC-84. FLOWRATE: 2 GPM/CU. FT.

FIG. 37



REMOVAL OF TOTAL ORGANIC CARBON FROM FLUSHING LIQUOR WITH AMBERLITE-IRC-84

FEED: TOTAL ORGANIC CARBON AS C 1600 MG/L
FLOWRATE: 2 GPM / CU. FT.

FIG. 38

Free Mineral Acidity (FMA) and pH of Amberlite 200 and Amberlite IRC-84 Effluent

The FMA and the pH of Amberlite 200 effluent remained constant for most of the exhaustion run. The FMA was 520 to 540 mg/l with a pH of 2.3 to 2.5. When the resin became exhausted, there was no FMA and the pH gradually increased to 6 or above. The Amberlite IRC-84 effluent showed 300 mg/l FMA at the start of the run, and then gradually decreased to 30 at 16 bed volumes. (See Figure 39) The pH of the IRC-84 effluent changed according to the FMA in the effluent. It started at 2.6 and then was increased to 4.9 at zero FMA. Results are shown on Figure 39.

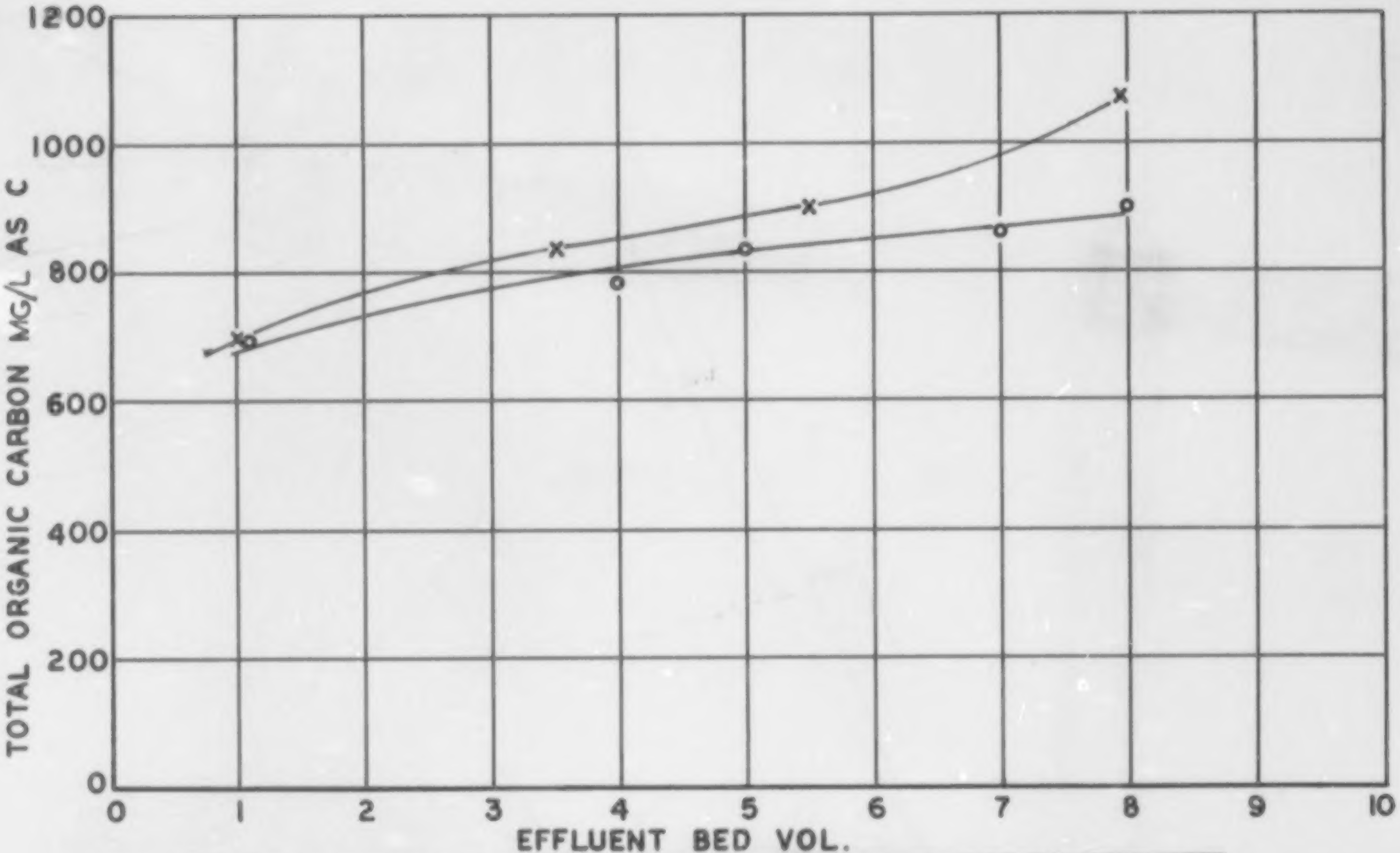
Phenol Reduction of Flushing Liquor with Ion Exchange Resins Amberlite 200 and Amberlite IRC-84

The phenol reduction with Amberlite 200 is shown on Figure 40 and that of Amberlite IRC-84 is shown on Figure 41. In both cases the phenol removal is small and it took place at the start of the exhaustion cycle. Some of the adsorbed phenol was released later in the run showing larger readings than in the flushing liquor feed.

Discussion of Results

Although Amberlite 200 shows a good ammonia removal, the operating capacity appears too small to make it economically feasible for flushing liquor treatment.

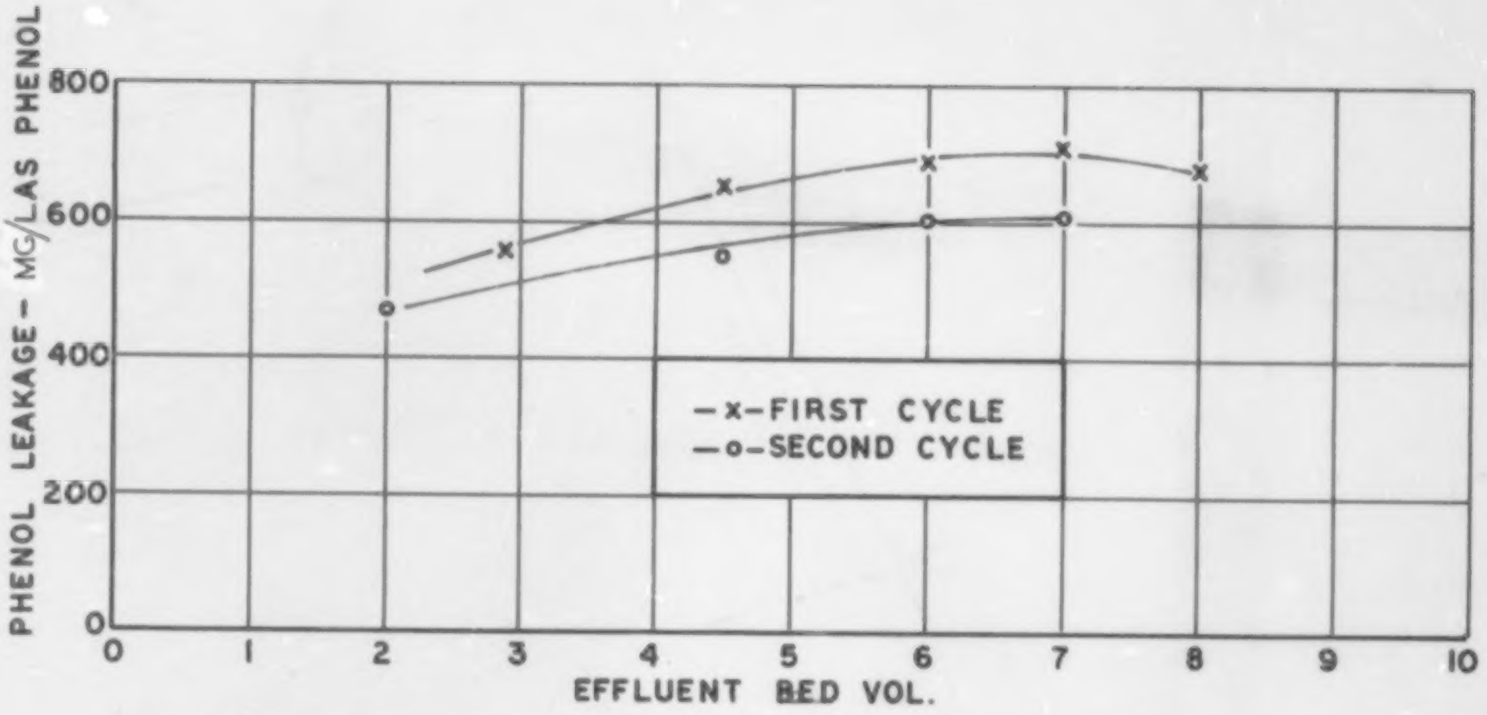
Amberlite IRC-84 showed a larger ammonia removal than Amberlite 200. The overall ammonia leakage of Amberlite IRC-84, however, was about 50% larger than with Amberlite 200. It showed 200 to 250 mg/l ammonia leakage, but the service run was twice as long as with Amberlite 200. Although the Amberlite IRC-84 has about 60% more capacity for ammonia, it is not recommended for the ammonia removal work because of its larger ammonia leakage and its swelling characteristics. Amberlite IRC-84 swelled about 20% during the exhaustion run. The excessive swelling may cause pressure drop and operating problems.



REMOVAL OF TOTAL ORGANIC CARBON
FROM FLUSHING LIQUOR WITH AMBERLITE 200

FEED: TOTAL ORGANIC CARBON AS C 1600 MG/L
FLOWRATE: 2 GPM/CU. FT.

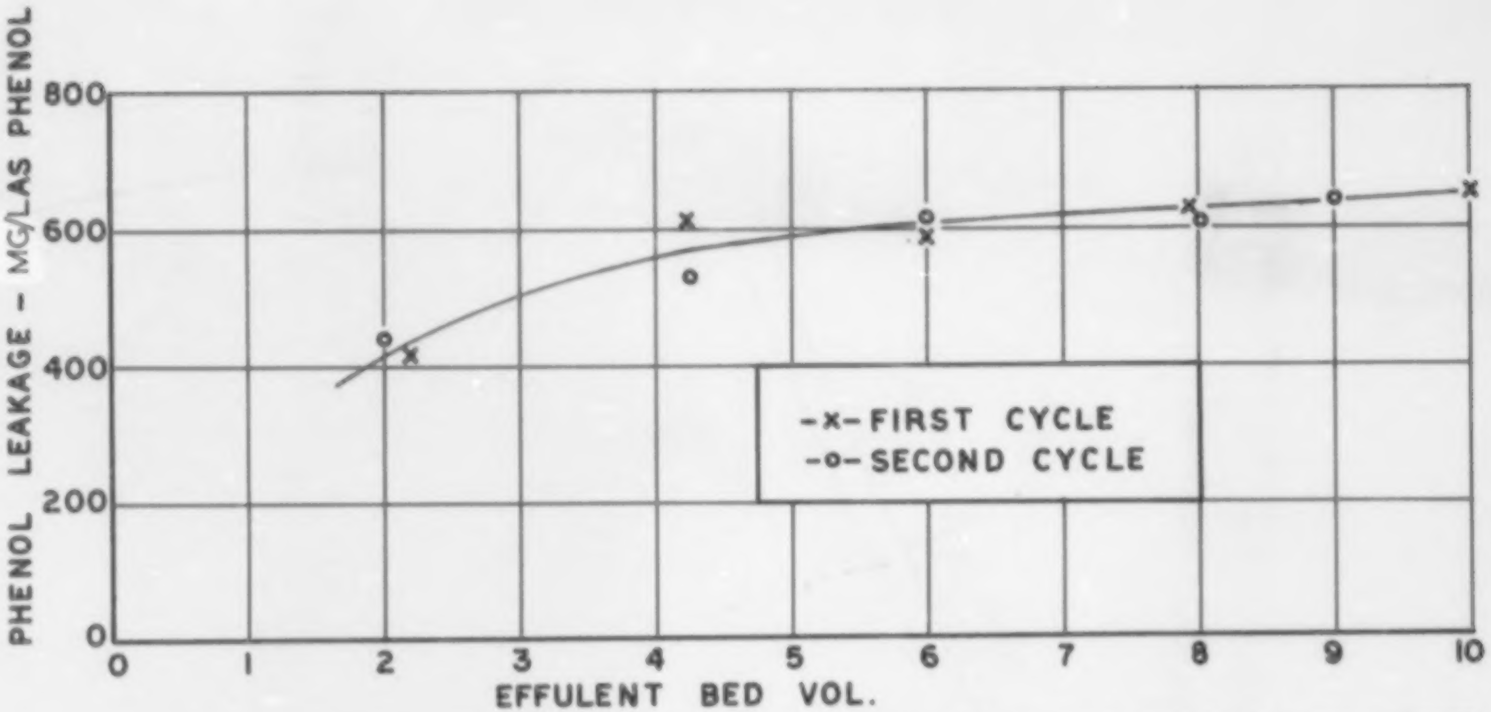
FIG. 39



REMOVAL OF PHENOL FROM FLUSHING LIQUOR
BY ADSORPTION WITH AMBERLITE 200 H FORM

FEED: TOTAL PHENOLS AS PHENOL 640 MG/L
FLOWRATE: 2 GPM/CU. FT.

FIG.. 40



REMOVAL OF PHENOLS FROM FLUSHING LIQUOR BY ADSORPTION WITH AMBERLITE IRC-84

FEED: TOTAL PHENOLS AS PHENOL 640 MG/L
FLOWRATE: 2 GPM/CU. FT.

FIG. 41

Ion exchange process even with somewhat idealized operating conditions requires a high degree of operator attention and frequent regeneration and is thus relatively expensive to operate. This fact coupled with the added operational problems indicated in the screening tests reduced capacity, and the failure to remove cyanide or phenols nearly preclude consideration of the cation resins tested for treatment of the coke plant wastes.

FILTRATION AND SEDIMENTATION COAL WASHERY

Laboratory - Bench Scale Tests

The coal washery operates only during the second shift. It discharges approximately 360,000 gallons each day at a rate of 600 gallons per minute. The discharge contains 4 to 12% suspended solids and is sent directly to the coke plant sewer with minimal treatment in a small settling sump.

A 1962 survey of bituminous coal and lignite preparation plants showed that 82% of the combined recycle water from all plants was treated in thickeners or settling ponds[1]. Use of thickeners provides greater recycle which results in less discharge flow. Actually, the only discharge of most plants is in the form of a highly concentrated solids stream which is sent to the sludge lagoons for disposal or dewatered by vacuum filters or centrifuges. These approaches appeared to be the most logical avenues to take for study of the CF&I coal washery. Settling tests were conducted using 500 ml. samples of the coal washery waste to determine per cent suspended solids in the supernate vs. settling time. Since unaided settling of the present coal washery effluent does not yield a sufficiently treated supernate (Figure 42), organic and inorganic coagulants were tested for their ability to increase settleability and clarification of the effluent.

Screening Tests

Various combinations and dosages of reagents were added to 500 ml. samples of coal washery effluent in order to determine the requirements for clarification and concentration of this waste (Tables 50 and 51). Rapid settling of sludge and good clarification of the supernate were the two criteria used to evaluate a reagent or combination of reagents.

Concentrations of 100 mg/l alum, a cationic coagulent, in combination with 5 mg/l Nalcolyte 673 (an anionic coagulent) produced good clarity and settleability.

Settling Tests

Interfacial settling rates were all determined on 500 ml. portions of test samples in 500 ml. graduated cylinders having a column height of 11 inches at the 500 ml. volume mark. The procedure for the settling tests was as follows: The first reagent was added to 500 ml. of sample and the sample was slowly inverted four

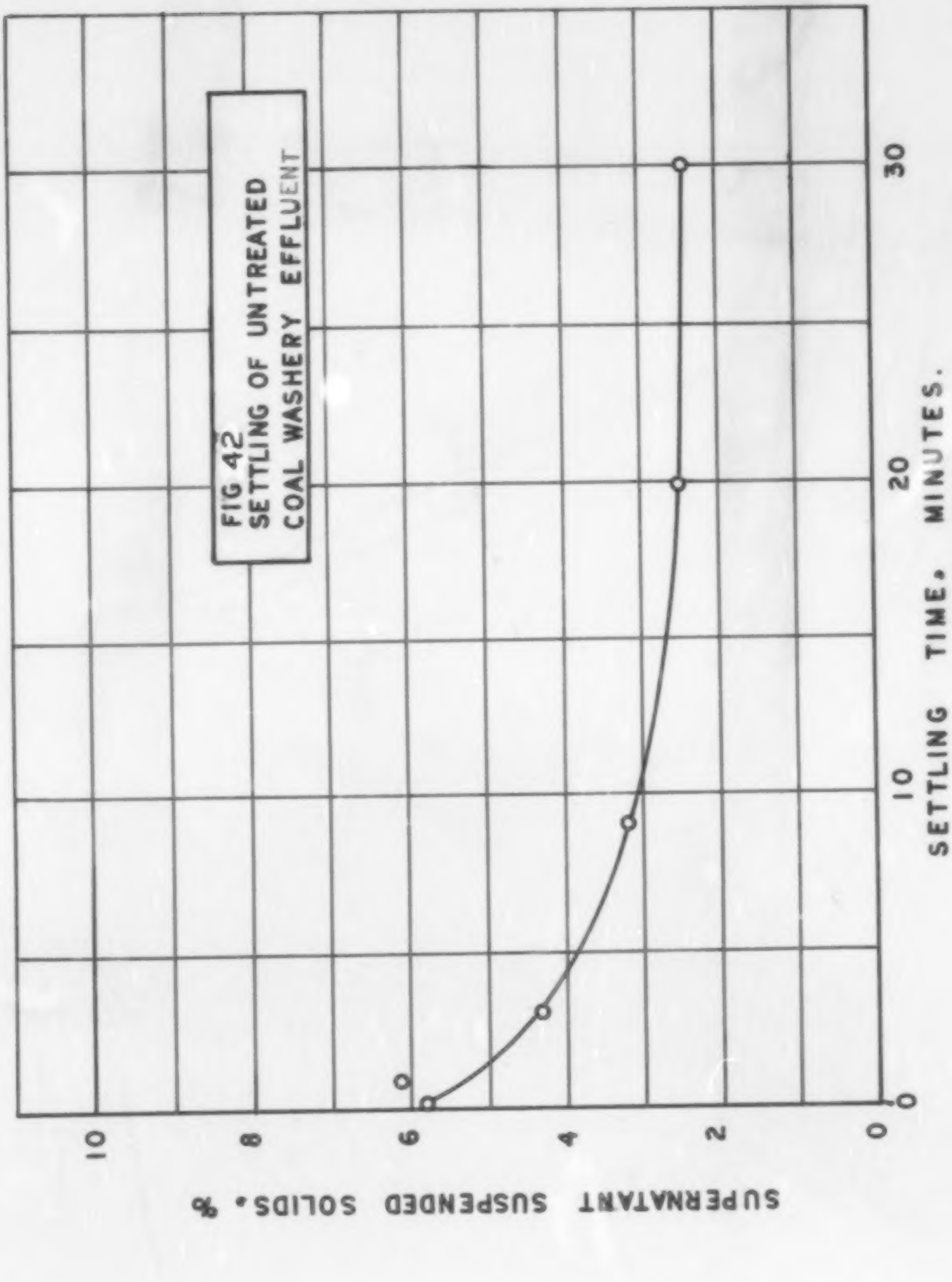


FIG 42
SETTLING OF UNTREATED
COAL WASHERY EFFLUENT

TABLE 50

REAGENT SCREENING FOR CLARIFICATION OF COAL WASHERY EFFLUENT

Reagent I Dosage mg/l	Reagent II Dosage mg/l
Alum- 40	670-2
40	8
Alum-100	672-2
100	4
Alum- 20	673-3
20	5
30	2
40	2
40	4 S
40	6 S
50	1
50	2
70	2
80	4
80	7 C
100	1
100	2 C
100	3 C
100	5 CS
100	6 CS
150	2 C
200	4 C

C = Fair to good clarity
S = Sludge interface settled to less than 60% of initial height within 10 minutes
No comment indicates poor clarity and settleability.
Three separate coal washery effluent grab samples were used in the the screening tests.
Alum - An acidic cationic inorganic electrolyte
Nalco 670 - A neutral nonionic organic polyelectrolyte
Nalco 672 - A neutral anionic organic polyelectrolyte
Nalco 673 - A basic anionic organic polyelectrolyte

TABLE 51

REAGENT SCREENING FOR CLARIFICATION OF COAL WASHERY EFFLUENT

Reagent I Dosage mg/1	Reagent II Dosage mg/1	Reagent I Dosage mg/1	Reagent II Dosage mg/1
Alum - 200	673- 4 C	603- 3	673- 1
200	6 CS	4	2
		4	5
Alum - 100	674- 2	5	1
100	4	5	2 S
		10	4 S
Alum - 40	675- 1	10	5 S
40	5 S	15	4 S
		20	4 S
Lime - 20	673- 2 S	603- 2	675- 2
50	2 S	10	1
100	2 S	20	2
200	2 S	40	2
Sodium Aluminate-50	675- 1	605- 2	2 S
50	3	5	1
50	5	6	2 S
50	9	10	2 S
50	13	20	2 S
150	2 S	20	4 S
603 - 5	670- 1	673- 2	603- 5 S
603 - 1	673- 1	673- 2	--
1	5		

C = Fair to good clarity
S = Sludge interface settled to 60% of initial height in 10 min.
No comment indicates poor clarity and settleability.
Three separate coal washery effluent grab samples.
Alum - An acidic cationic inorganic electrolyte
Lime - A basic cationic inorganic electrolyte
Sodium Aluminate - A basic cationic inorganic electrolyte
Nalcolyte 603 - A basic cationic organic polyelectrolyte
Nalcolyte 605 - A basic cationic organic polyelectrolyte
Nalco 670 - A neutral nonionic organic polyelectrolyte
Nalco 672 - A neutral anionic organic polyelectrolyte
Nalco 673 - A basic anionic organic polyelectrolyte
Nalco 674 - A basic anionic organic polyelectrolyte
Nalco 675 - A basic anionic organic polyelectrolyte

times to mix. The second reagent, if used, was then added and again the cylinder was slowly inverted four times. Treated samples were then allowed to stand undisturbed while the height of the sludge interface was recorded with time. Supernatant samples were withdrawn 2" below the surface and analyzed for suspended solids. Settling tests were run on samples treated with (a) an anionic polymer, Nacholyte 673 (Figure 43a); (b) a basic cationic polymer, Nalcolyte 603, followed by 673 (Figure 43b); (c) sulfuric acid followed by 673 (Figure 44), and (d) waste hydrochloric acid pickle liquor followed by 673 (Figures 45-47). The combination of 603 and 673 has been commonly used in other coal preparation plants but did not achieve a result as good as 673 alone for the CF&I waste (figure 43b).

Visually, the supernates of samples treated with 603 followed by 673 looked much better than samples treated with 673 alone. Suspended solids results did not verify this (Figure 43). Alum, which is an acidic cationic coagulant, did help to achieve supernate clarity in combination with 673. However, large alum dosages were required which consequently increased the sludge volume and decreased the settling rate (Figure 49).

Heavy dosages of alum lower the pH. To test the effect of pH on clarification, washery effluent ($pH > 8.0$) was first neutralized with acid prior to coagulant addition. Addition of sulfuric acid was found to destabilize the colloidal system and produce a sparkling clear supernate. However, this additive did not promote rapid settling. Addition of 673 to a neutralized sample produced a rapidly settling sludge (Figure 44) with a clear supernate containing little or no suspended solids. Thus, neutralization followed by coagulation with an anionic polymer appeared to be the answer for clarification of this waste.

Hydrochloric acid pickle liquor is available at this plant as a waste product needing treatment. Considering this as a cheap source of acid, tests were conducted to determine whether this liquor could destabilize the colloidal system without interferring with the settleability of the waste. It was found that this liquor would also work well in breaking the colloidal system (Figures 51-52). Supernatant turbidity was measured using a phototester to compare the sample turbidity against turbidity of distilled water. A volume ratio of less than 1 part pickle liquor per 1,000 parts effluent (0.1% by vol.), was found suitable for destabilization. Higher ratios were found to leave significant residual iron in the supernate (Table 53) which would be undesirable from a corrosion standpoint. Furthermore,

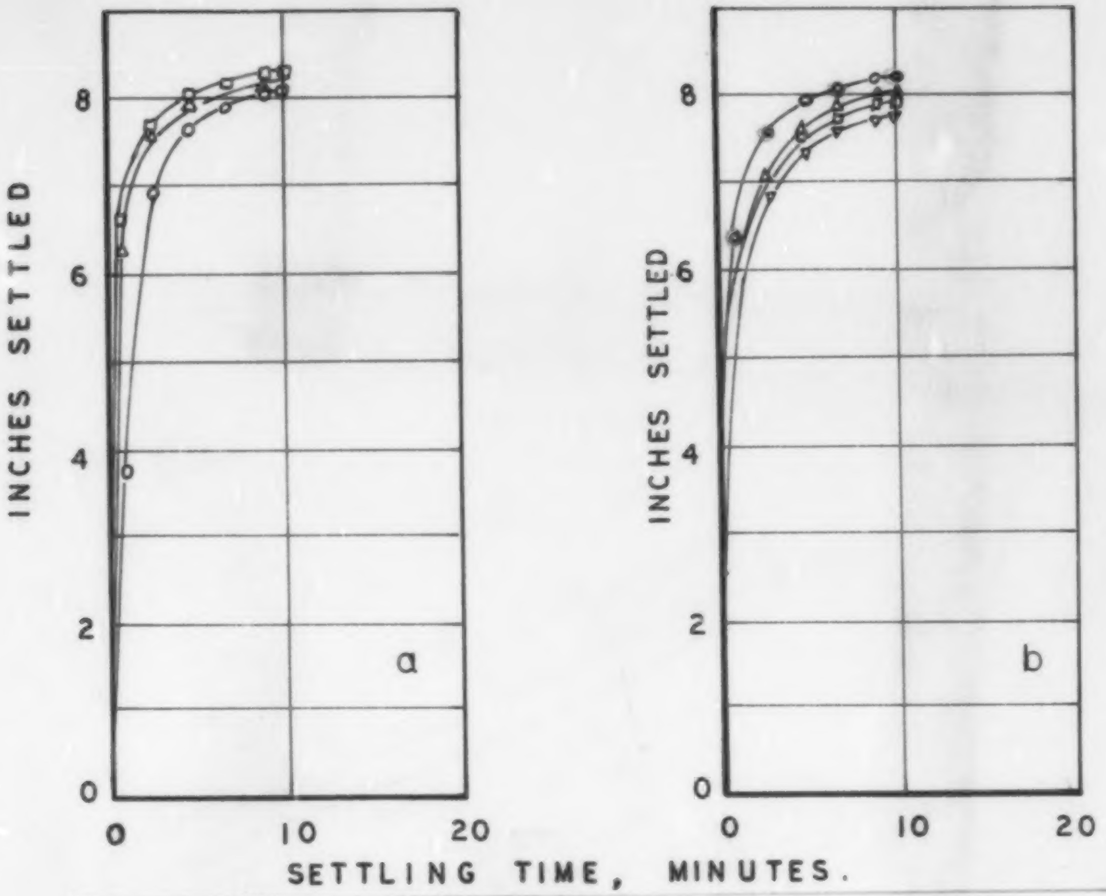


FIG.43
INTERFACIAL SETTLING RATE OF TREATED COAL WASH. EFFLUENT. 11 INCH SETTLING COLUMN

a. ANIONIC POLYMER ALONE

TEST	NALCO 673 MG/L
—○—	2
—△—	4
—□—	6

b. CATIONIC (NALCO 603) AND ANIONIC POLYMER, (NALCO 673)

TEST	603 MG/L	673 MG/L	SUPERNATE SUSP. SOL. 10 MIN.
—○—	0	4	754 MG/L
—△—	5	4	—
—□—	10	4	—
—▽—	20	4	3226 MG/L

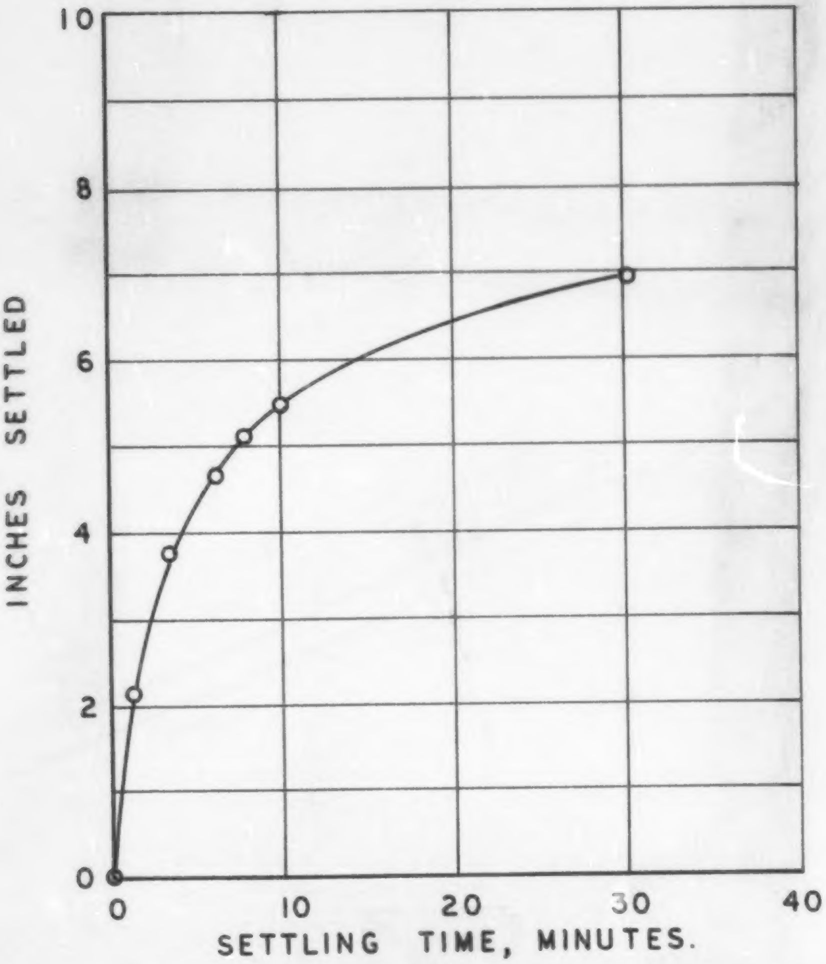


FIG.44

INTERFACIAL SETTLING OF COAL WASH.
EFFL. pH ADJUSTED TO 7.0 WITH H_2SO_4.
6MG/LNALCO 673 ADDED.
11" DEEP SETTLING COLUMN.

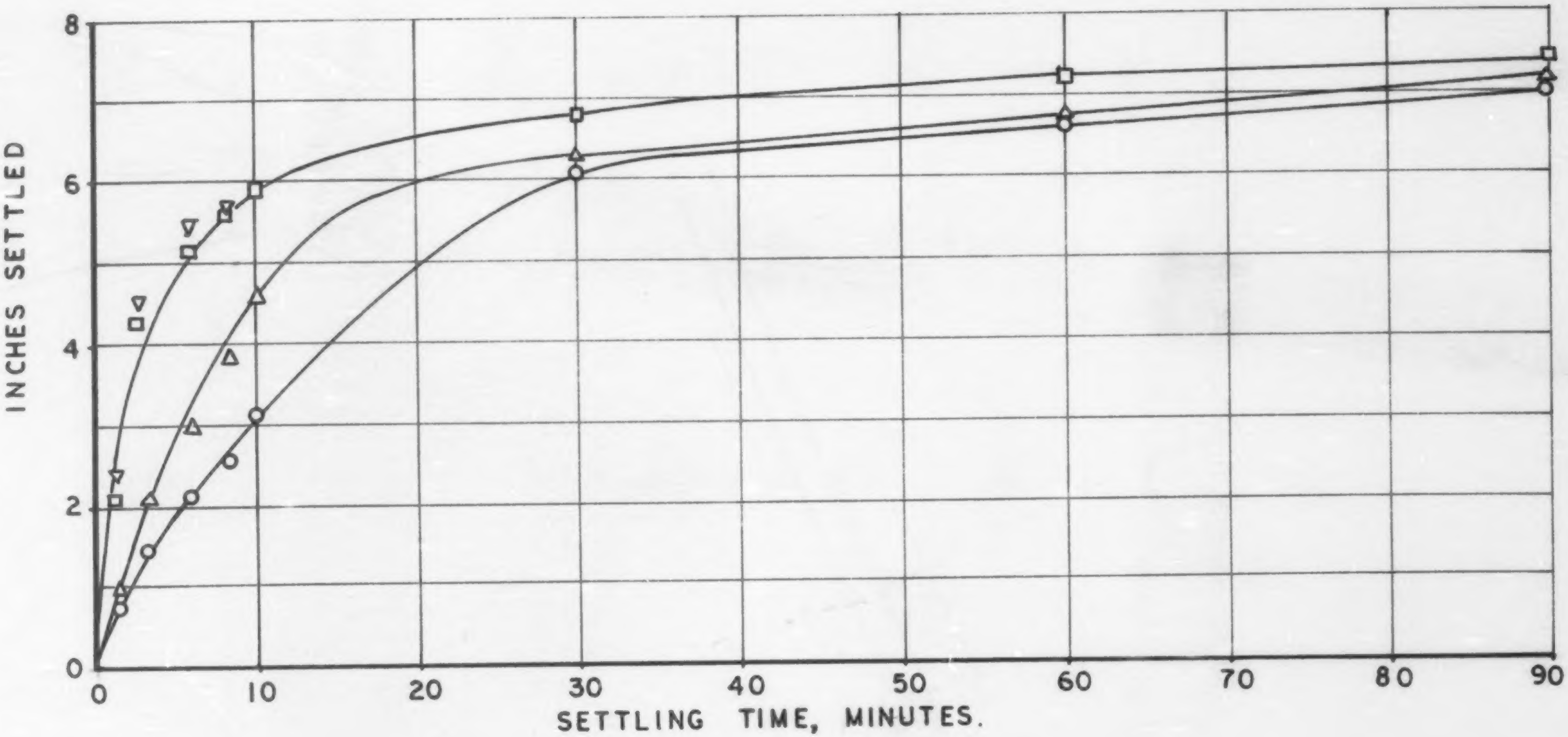


FIG.45
INTERFACIAL SETTLING RATE OF TREATED COAL WASHERY EFLUENT. 11 INCH COLUMN.
ALL TESTS WITH .02 % HCl PICKLE LIQUOR (BY VOLUME)

TEST	NALCO 673 MG/L	SUPERNATE PHOTOTESTER READING AT 100 MIN. SETTL.
○	3	68
△	4	54
□	5	55
▽	6	55

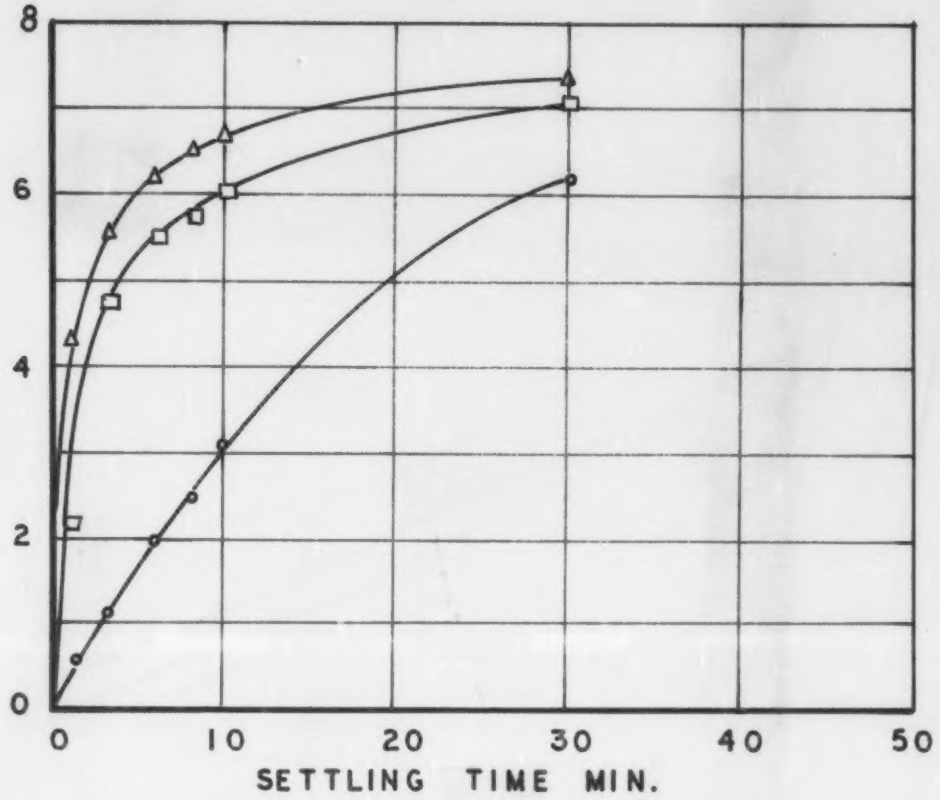


FIG.46
INTERFACIAL SETTLING RATE OF TREATED COAL WASHERY EFFL. 11 INCH SETTLING COLUMN.

0.06 % BY VOL. HCl PICKLE LIQUOR AND NALCO 673

TEST	NALCO 673 MG/L
-○-	4
-□-	6
-△-	8

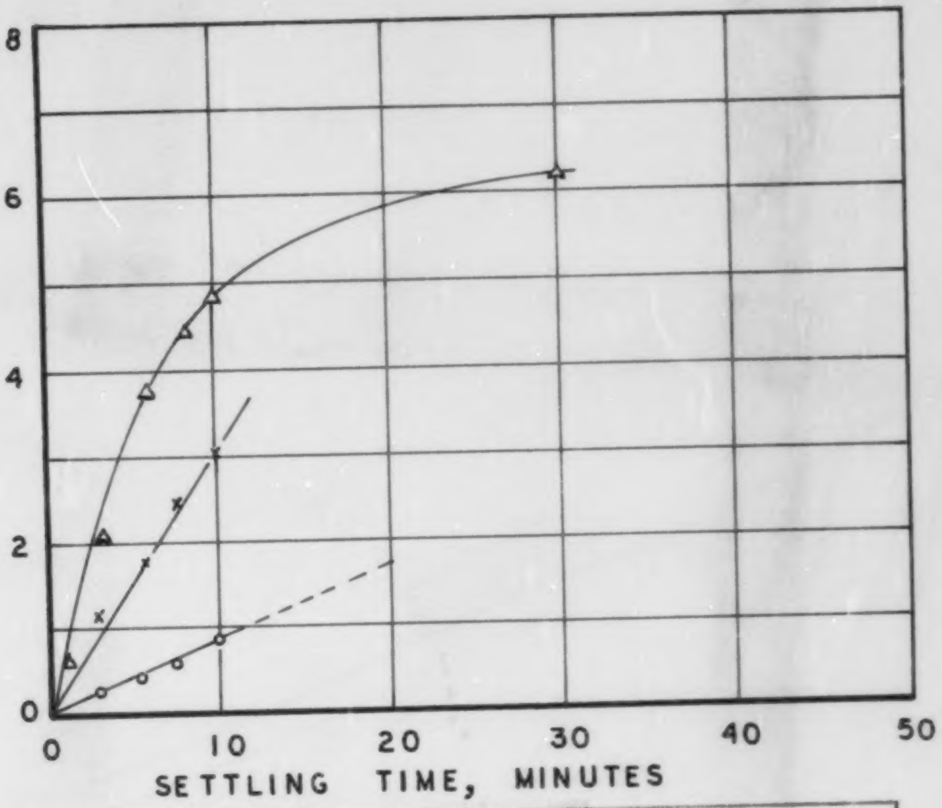


FIG.47
INTERFACIAL SETTLING RATE OF TREATED COAL WASHERY EFFLUENT. 11 INCH SET. COLUMN.

0.10 % BY VOL. HCl PICKLE LIQUOR AND POL. EL.

TEST	NALCO 673 MG/L
–○–	2
–×–	4
–△–	6

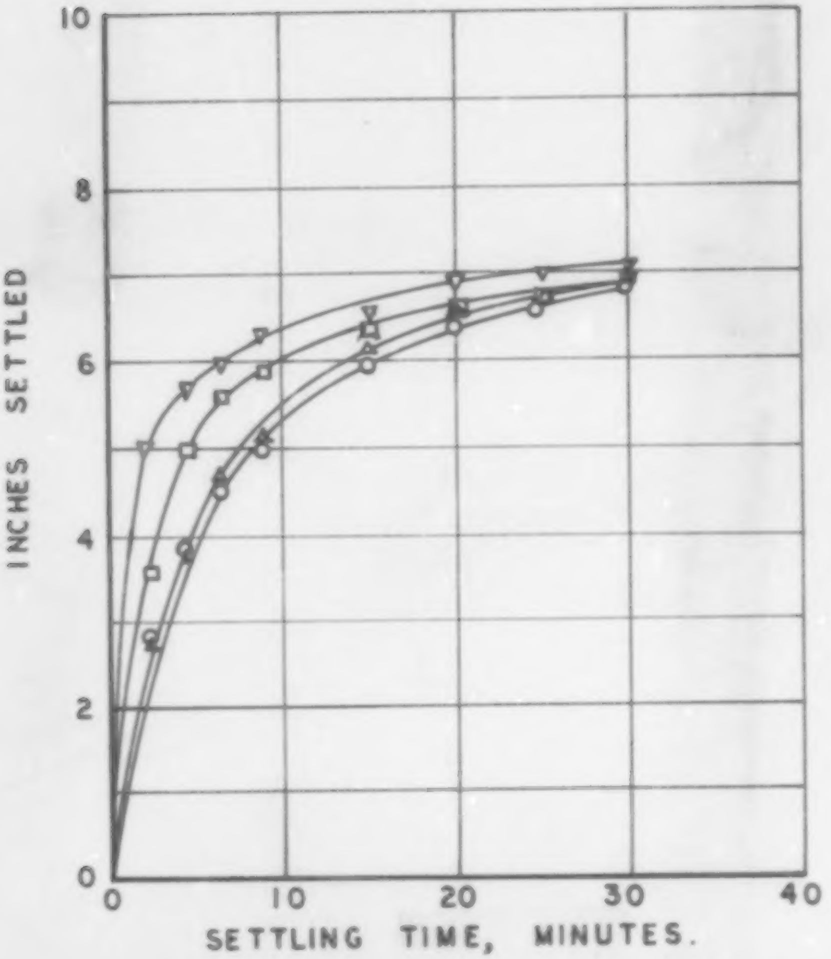


FIG. 48
INTERFACIAL SETTLING RATE OF TREATED COAL W. EFFL. ALL TESTS WITH 10 MG/L NALCO 603. 11 INCH SETTLING COLUMN.

TREATMENT AND CLARIFICATION

	NALCO 603 MG/L	SUPERNATE AT 10 MIN.	SUSP. SOL. MG/L AT 30 MIN.
—○—	.5	—	22,200
—△—	1.0	—	—
—□—	2.0	1306	—
—▽—	4.0	2427	3049

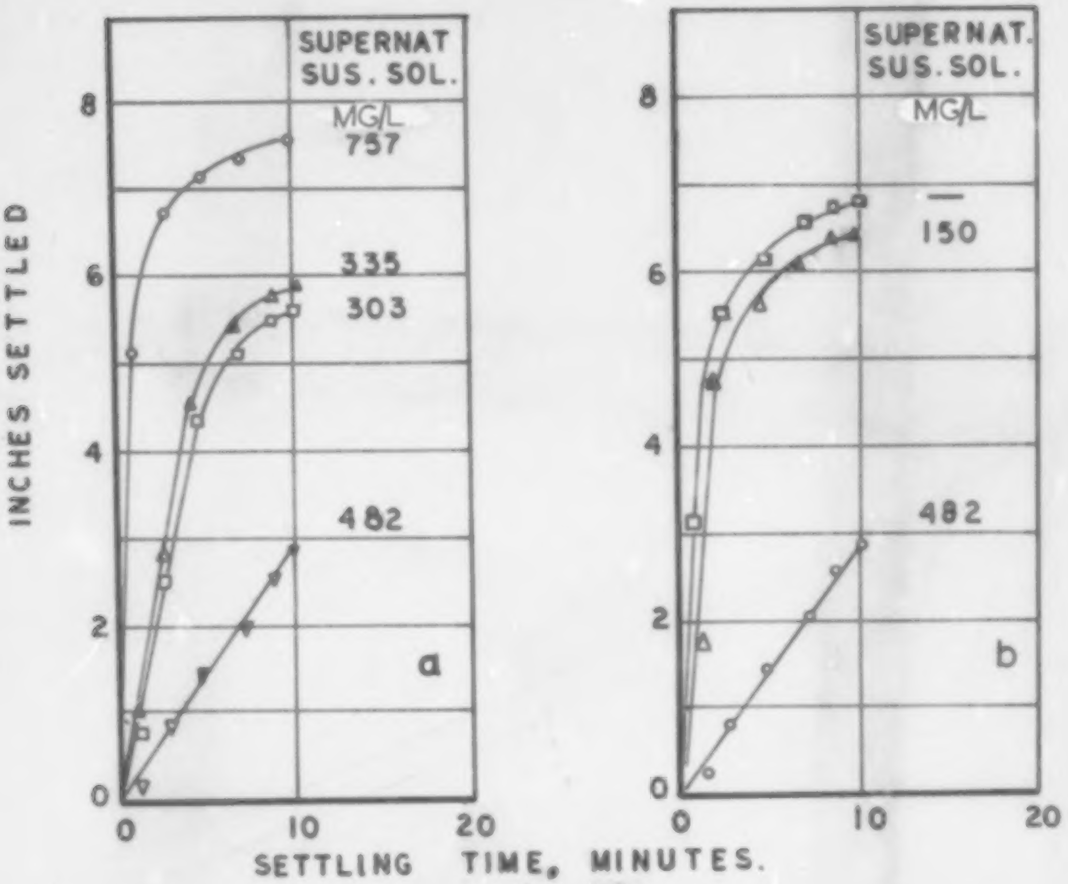


FIG. 49
INTERFACIAL SETTLING RATE OF TREATED COAL WASHERY EFFLUENT. 11 INCH SETTLINC COLUMN ALUM AND ANIONIC POLYELECTROLYTE (NALCO 673)

	TEST	ALUM MG/L	673 MG/L	pH
a.	-○-	50	4	7.8
	-▲-	100	4	7.6
	-□-	150	4	7.4
	-▼-	200	4	7.3
b.	-○-	200	4	—
	-△-	200	6	—
	-□-	200	8	—

TABLE 52

ANALYSIS OF SAMPLES FROM COAL WAHSERY

Sample Source	Combined Jig Recycle	South Jig Recycle	North Jig Recycle	Combined Jig Recycle	Combined Jig Recycle
Type	Grab	Grab	Grab	Grab	Grab
Date	2/19/69	2/20/69	2/20/69	3/24/69	10/29/69
Total Solids (%)	5.83	7.86	9.95	9.05	13.5
Percent Ash* (%)	41.4	37.0	35.5	37.4	38.0
pH	8.3	-	-	-	-
Sulfur* (%)*	-	-	-	1.1	-

Sample Source	Tailings Ditch	Tailings Ditch	Tailings Ditch	Tailings Ditch	Tailings Ditch
Type	Grab	Grab	Grab	Grab	Grab
Date	2/19/69	2/19/69	3/25/69	3/26/69	10/29/69
Total Solids (%)	5.82	4.65	10.5	7.4	12.0
Percent Ash* (%)	52.6	53.7	42.9	57.5	42.2
pH	8.1	8.5	-	-	-
Temperature (°C)	9°C	9°C	-	-	-
Sulfur* (%)*	-	-	1.4	-	-

*Dry Basis

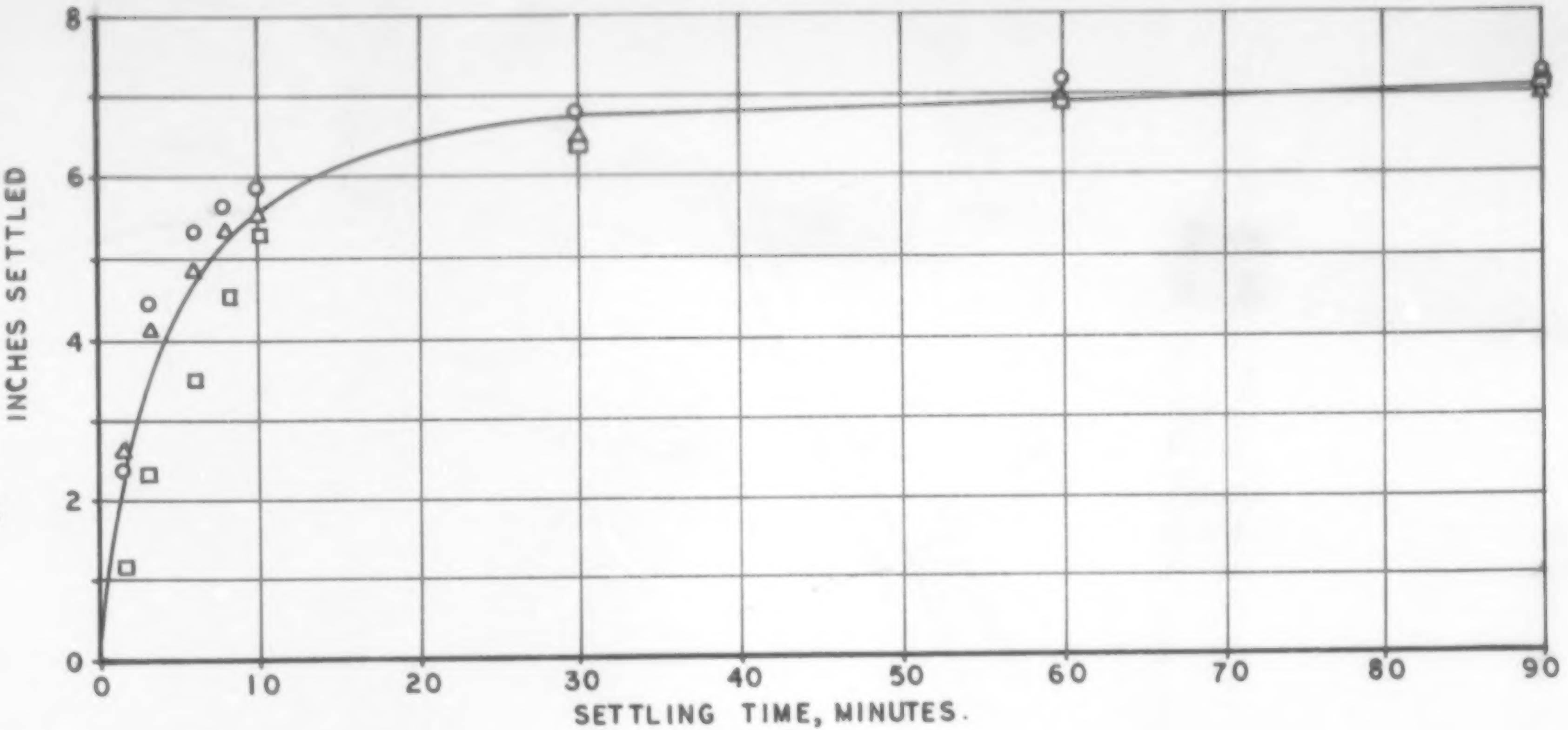


FIG. 50

INTERFACIAL SETTLING RATE OF TREATED COAL WASHERY EFFLUENT. 11 INCH COLUMN.

TEST	PICKLE LIQ. VOL. %	SUPERNATE PHOTOTESTER READING AT 100 MIN. SETTL.
—○—	.02	55
—□—	.03	22
—△—	.04	8

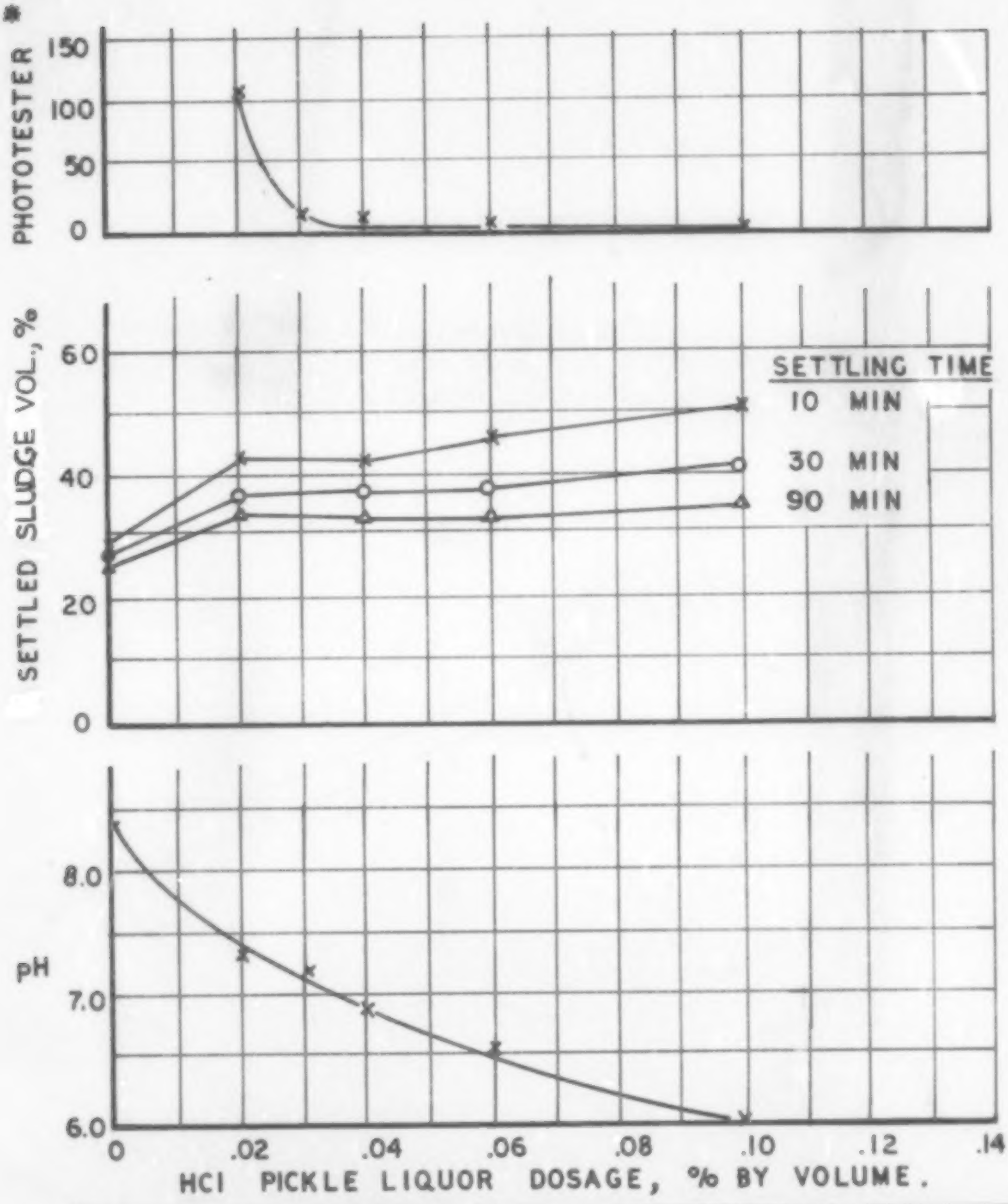


FIG. 51

TREATMENT OF COAL WASHERY EFFL. WITH HCl PICKLE LIQUOR AND 6 MG/L NALCO 673.

*READING ON SUPERNATANT AFTER 20 MIN. SETTLING.

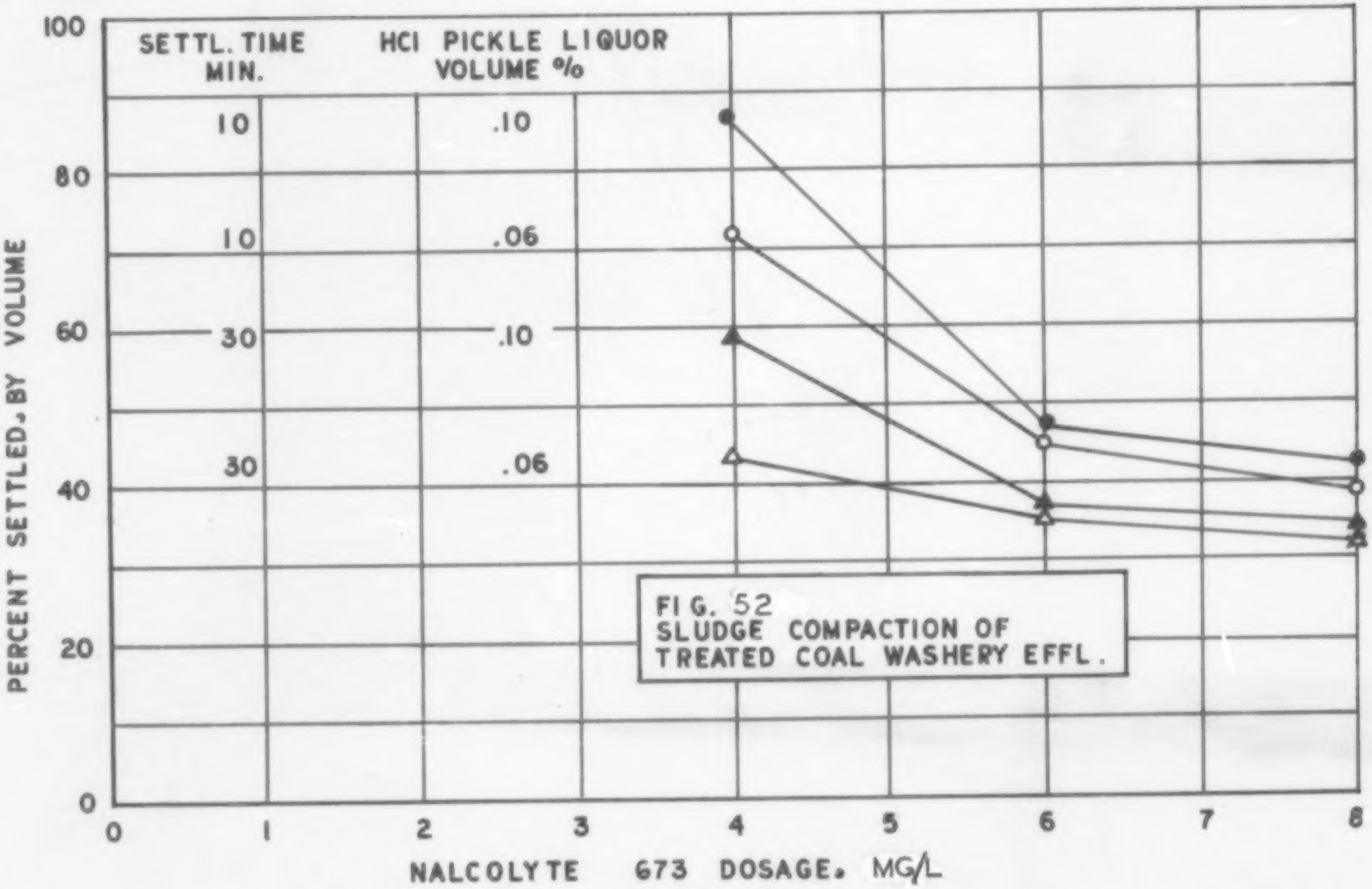


FIG. 52
SLUDGE COMPACTION OF
TREATED COAL WASHERY EFFL.

TABLE 53

SUPERNATANT IRON CONCENTRATION OF HCl - PICKLE LIQUOR TREATED COAL WASHERY EFFLUENT WITH AND WITHOUT LIME ADDITION

Reagent I	Reagent II	Reagent III	Readings After 10 Minutes Settling		
HCl-Pickle Liquor (Volume %)	Nalcolyte 673 (mg/1)	Lime (mg/1)	Settled Sludge Volume (ml)	Supernatant Phototester Reading	Supernatant Iron Concentration (mg/1)
.04	6	0	205	8	0.7
.08	6	0	220	4	5.3
.14	6	0	300	4	10.0
.08	6	0	220	4	5.3
.08	6	20	-	-	3.6
.08	6	100	220	4	0.5
.08	6	200	195	7	0.1

a ratio of 6 parts HCl pickle liquor per 100 parts washery effluent used in the screening tests resulted in gasification as well as floatation of some solids. Excessive dosages of lime were required to reduce the supernatant iron to satisfactory levels.

Reduction of the effluent pH with reagent grade sulfuric acid and waste hydrochloric acid pickle liquor is shown in Figure 53. It can be seen from this figure that a 1:10 dilution of pickle liquor had about the same neutralizing capability as 1N H_2SO_4.

The optimum pH for good clarification is about pH 7 (Figure 51). The results of these tests indicated that the settled sludge volume decreased with increasing acid dosage (decreasing pH). The initial rate of sludge compaction decreased with increasing acid dosage (decreasing pH).

In addition, it is apparent that an optimum dosage of anionic polymer exists (Figure 52). The cost of the anionic polymer makes it necessary to determine and closely control this optimum dosage. Control of the acid dosage does not greatly effect the system operating cost since it is available as an excess waste product.

Treatment with Acid and Polyelectrolyte

The treatment of the coal washery wastes was pursued further in order to gain additional information on acid requirements, sludge settling rates, thickener size requirement and methods of sludge dewatering. The method of treatment selected will depend on the settling rate of the wastes. If the clarified waste settles at a reasonable rate, consideration may be given to a clarifier-thickener located next to the coal washery. If, on the other hand, the settling rates are found to be extremely low, in-ponding of the entire waste from the washery may be necessary.

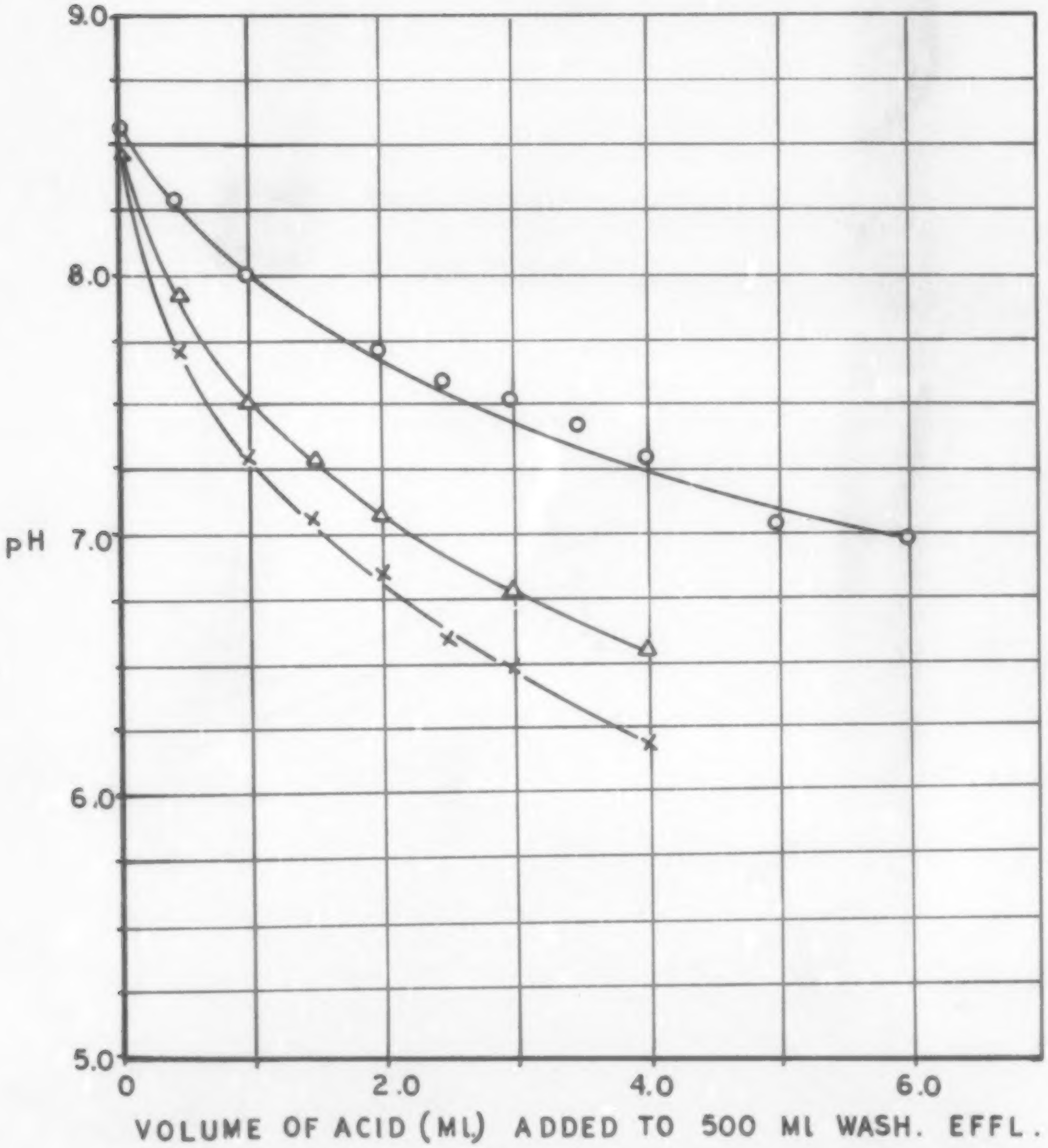


FIG. 53

ACID TITRATION OF COAL WASHERY EFFLUENT

TEST	ACID	
—×—	1 N H_2SO_4	
—△—	HCl PICKLE LIQUOR	1/10 DILUTION
—○—	HCl PICKLE LIQUOR	1/20 DILUTION

Acid Requirements

Since the mill has spent H_2SO_4 pickle acid available, use of this waste for pH adjustment was considered. The plant also has a galvanizing line in operation where hydrochloric acid is discharged periodically.

Samples of each acid were used for titration of the coal washery waste. Figure 54 shows titration curves of washery wastes with sulfuric and hydrochloric acids. The figure shows that approximately 1 gallon of either acid is required per 1000 gallon sample to reach the pH range 6.0-7.0.

In tests conducted in the 55 gallon drums (see Table 54) 200 ml. (0.053.gal.) of the sulfuric acid pickle liquor was used in most instances to treat 52 gallons of waste. These results indicate that the daily amount of acid required to treat the total waste flow from the coal washery is approximately .1% of the washery flow by volume. In other words, if the flow is 500 gpm for 10 hours, the total acid consumption, assuming no drastic change in concentration of the spent pickle acid, will amount to 300 gallons of acid.

If spent hydrochloric acid from the galvanizing line is used, the daily consumption would be slightly less than 300 gpd as indicated in Figure 54.

Settling Tests

A series of tests was conducted using 500 milliliter cylinders in order to evaluate the performance of hydrochloric acid and sulfuric acid in clarifying the colloidal system. Some of the tests were conducted with Nalco 603, Nalco 673, and a combination of the two. Some tests were run with a high molecular weight anionic polyelectrolyte, Nalco D-2339. Results of these studies are shown in Figures 55 and 56. Substantial quantities of polyelectrolytes were required to establish reasonable settling rates. Nalco 673 together with Nalco 603 gave improved settling rates at increasing electrolyte loading. Table 55 summarizes the pertinent data obtained from the tests. Comparisons made indicate Nalco D-2339 is more efficient and less expensive than the combination of Nalco 603-673. The daily cost of the D-2339 is $36 for a 600 gpm flow, while the minimum Nalco 603-673 dosages require an expenditure of $47.85 per day. Calculations were made as follows:

The average measured flow at the Coke Plant Coal Washery is 600 gpm. The washery normally operates on a 10 hour per day 24 four hour per week schedule. Total daily flow would then be 360,000 gallons per day.

Total Daily Flow = GPM X Operating Time Minutes = Gallons/Day
Total Daily Flow = 600 X 600 = 360,000 Gallons/Day

Since one pound in 120,000 gallons is equal to 1 mg/l, it is necessary to feed three pounds of coagulent per day to obtain 1 mg/l in the washery flow.

$$\text{Coagulent Feed} = \frac{360,000}{120,000} = 3 \text{ lbs./day for 1 mg/l}$$

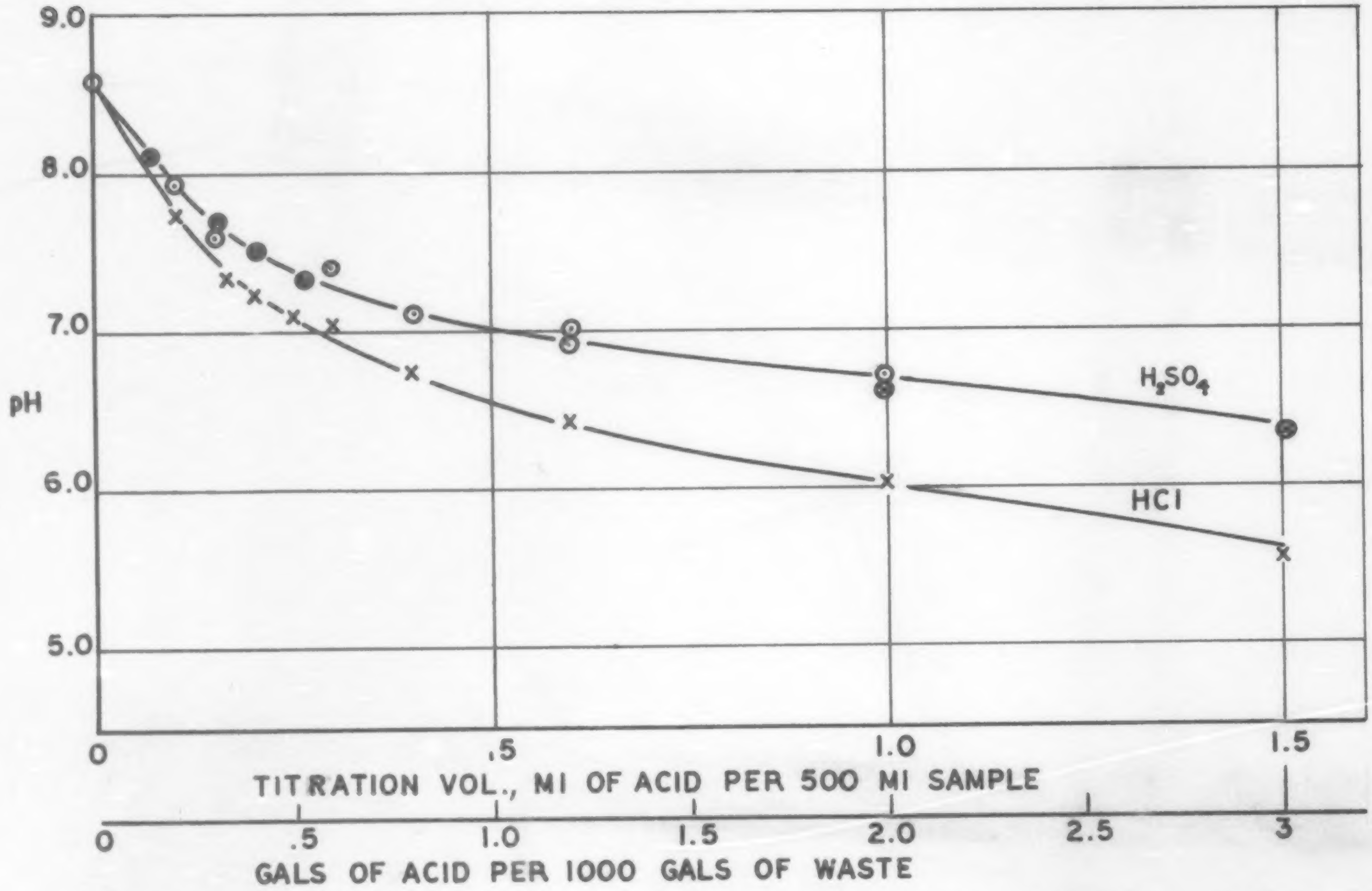
From the tests performed, a feed of 8 mg/l of Nalco D-2339 is needed for satisfactory settling in a clarifier. Thus a total of 24 pounds of Nalco D-2339 must be fed to the system each day. At $1.50/lb. x 24 lbs. required per day, daily chemical costs would be $36 per day.

Concentration Needed: 8 mg/l
Total feed/day = 3 x 8 = 24 lbs./day
Cost/day = 24 x $1.50/lb. = $36/day

In comparison, a combination feed of Nalco 603 (3 mg/l) and Nalco 673 (10 mg/l) is required for satisfactory settling. At these concentrations, a feed of 9 lbs. (Nalco 603) and 30 lbs. (Nalco 673) must be maintained. Daily chemical costs would be $0.25/lb. x 9 lbs. required per day or $2.25/day for Nalco 603 and $1.52/lb. x 30 lbs. required per day or $45.60/day or a total daily cost of $47.85/day.

Nalco 603 Concentration Needed = 3 mg/l
Total feed/day = 3 x 3 = 9 lbs./day
Cost/day = 9 x $0.25/lb. = $2.25/day

Nalco 673 Concentration Needed = 10 mg/l
Total feed/day = 3 x 10 = 30 lbs./day
Cost/day = 30 x $1.52/lb. = $45.60/day



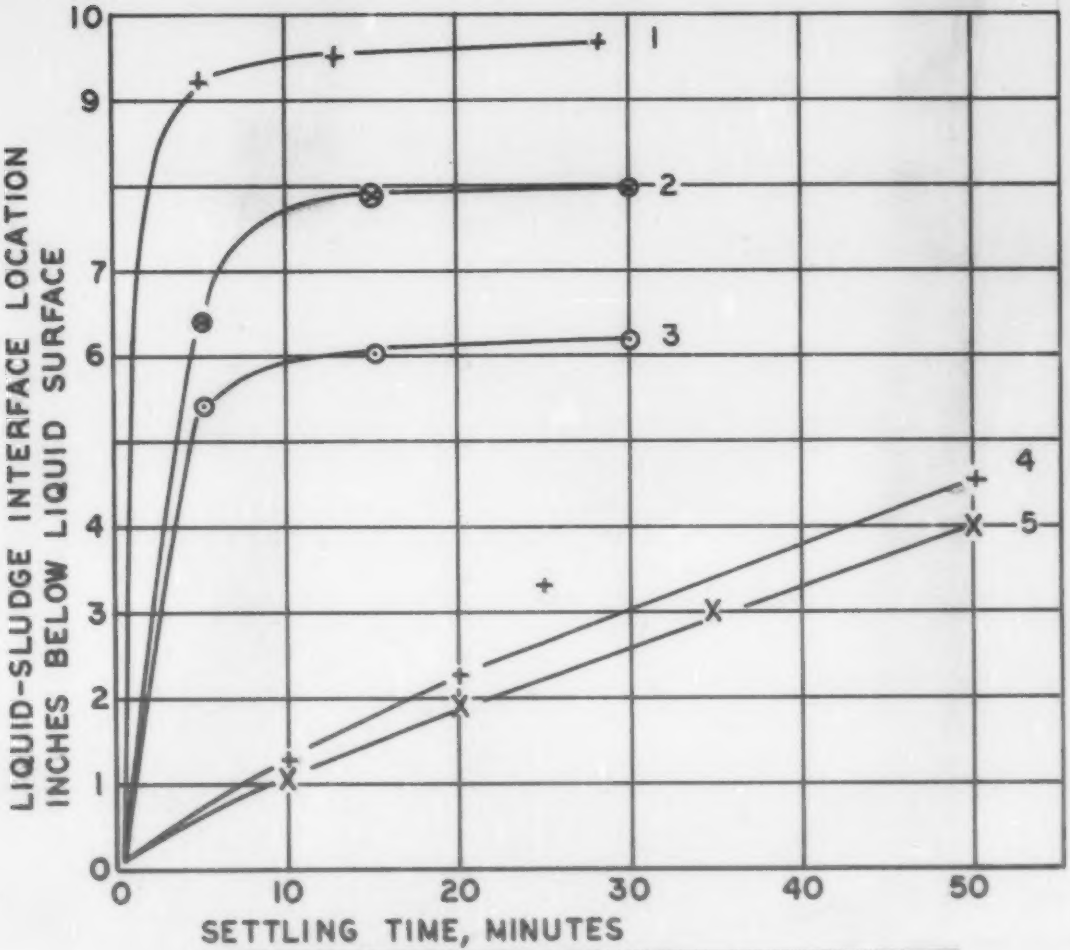
pH TITRATION CURVE FOR COAL WASHERY WASTE AUG. 4 1969

FIG. 54

TABLE 54

ANALYSES OF TREATED WATER FROM COAL WASHERY

Date	11-19	11-20	11-20	11-20	11-21
Size Sample, gal	52	52	52	52	52
Acid Used, H_2SO_4					
Volume Added, ml	300	200	200	200	200
Final pH	6.8	6.8	6.0	6.8	-
Nalco D-2339, mg/l	8	0	8	0	8
Settling Curve, Fig.	58	58	59	60	60
Curve No.	3	1	1	1	2
Settled Samples Analysis of Supernate					
pH (lab)	7.6	7.8	7.1	7.6	7.4
Total Suspended Solids mg/l	67	83	43	440	36
Total Hardness, $CaCO_3$ mg/l	66	50	110	86	96
Ca Hardness, $CaCO_3$ mg/l	49	36	81	62	71
P-Alkalinity, $CaCO_3$ mg/l	0	0	0	0	0
M-Alkalinity, $CaCO_3$ mg/l	261	256	166	238	212
Sulfate, Na_2SO_4 mg/l	490	445	408	425	408
TDS	760	690	550	620	610
Cyanide, CN mg/l	-	0.000	-	-	0.000
Phenol, C_6H_5OH mg/l	-	0.003	-	-	0,021
Langlier Index	-.13	-.38	-.60	-.06	-.26

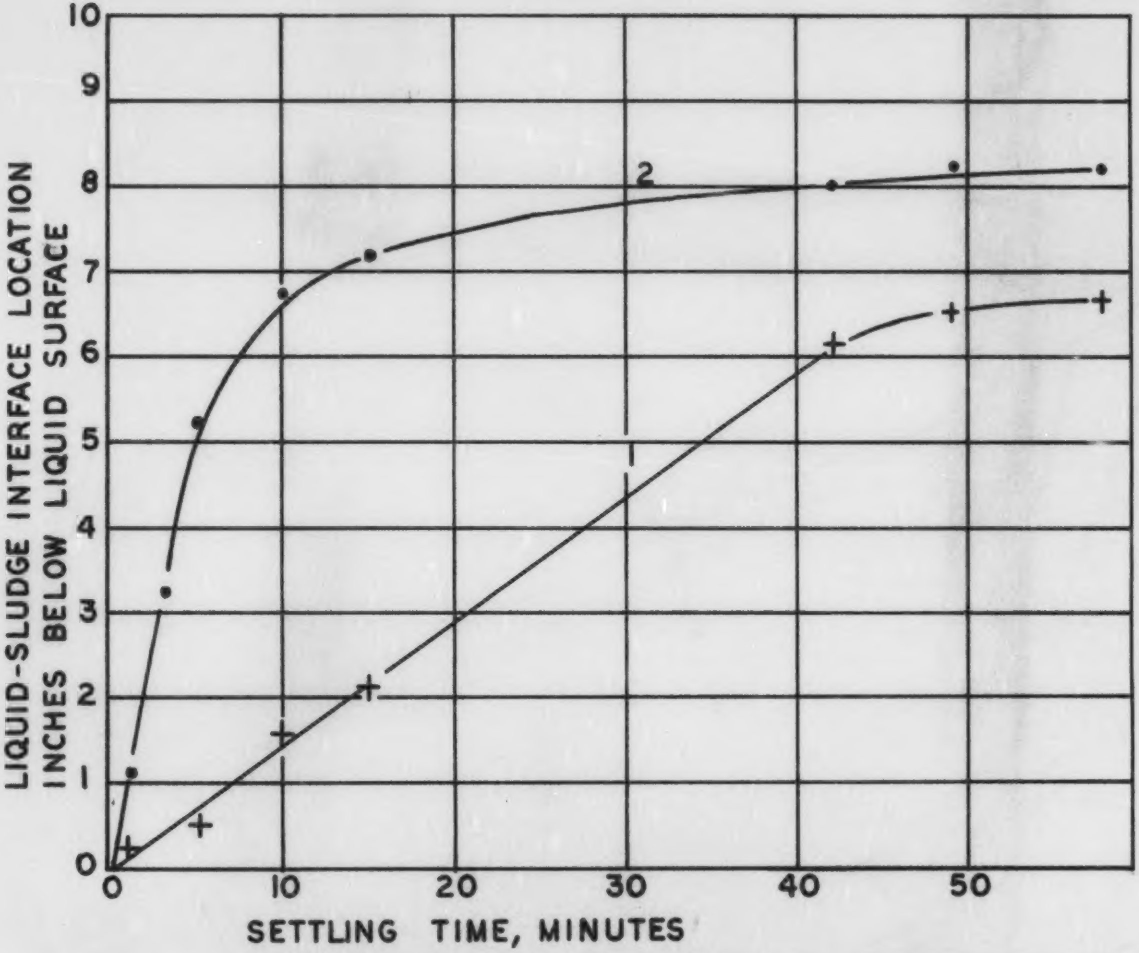


TYPE OF TREATMENT

CURVE NO.	ACID ADDED	FINAL pH	N-603 MG/L	N-673 MG/L
1	NONE	8.6	6	20
2	HCl	7.0	6	10
3	HCl	7.0	3	10
4	HCl	7.0	6	0
5	H_2SO_4	7.0	6	0

SETTLING CURVES 500 ML CYLINDERS

FIG. 55



CURVE NO.	TREATMENT ACID	FINAL pH	POLYELECTROLYTE
1	H_2SO_4	7.0	0 - 4 MG/L D-2339
2	H_2SO_4	7.0	8 MG/L D-2339

SETTLING CURVES. 500 MI CYLINDER

FIG. 56

TABLE 55

Comparison of Coagulents

Nalco 603-673 Combination vs. Nalco D-2339

	Inches Settled vs. Time (Min.)			Cost Chemical/ Lb.*			Chemical Added mg/l			Chemical Added Lbs./Day			
	10	20	30	603	673	D-2339	603	673	D-2339	603	673	D-2339	Cost/Day
Nalco 603-673													
Test 1	9.5	9.7	9.8	$0.25	$1.52	-	6	20	-	18	60	-	$95.70
Test 2	7.8	7.9	8.0	$0.25	$1.52	-	6	10	-	18	30	-	$50.10
Test 3	5.9	6.1	6.2	$0.25	$1.52	-	3	10	-	9	30	-	$47.85
Nalco D-2339													
Test 1	1.5	2.9	4.5	-	-	$1.50	-	-	4	-	-	12	$18.00
Test 2	6.6	7.4	7.8	-	-	$1.50	-	-	8	-	-	24	$36.00

*40,000 Lb. Carload Prices

Clarifier - Thickener Size

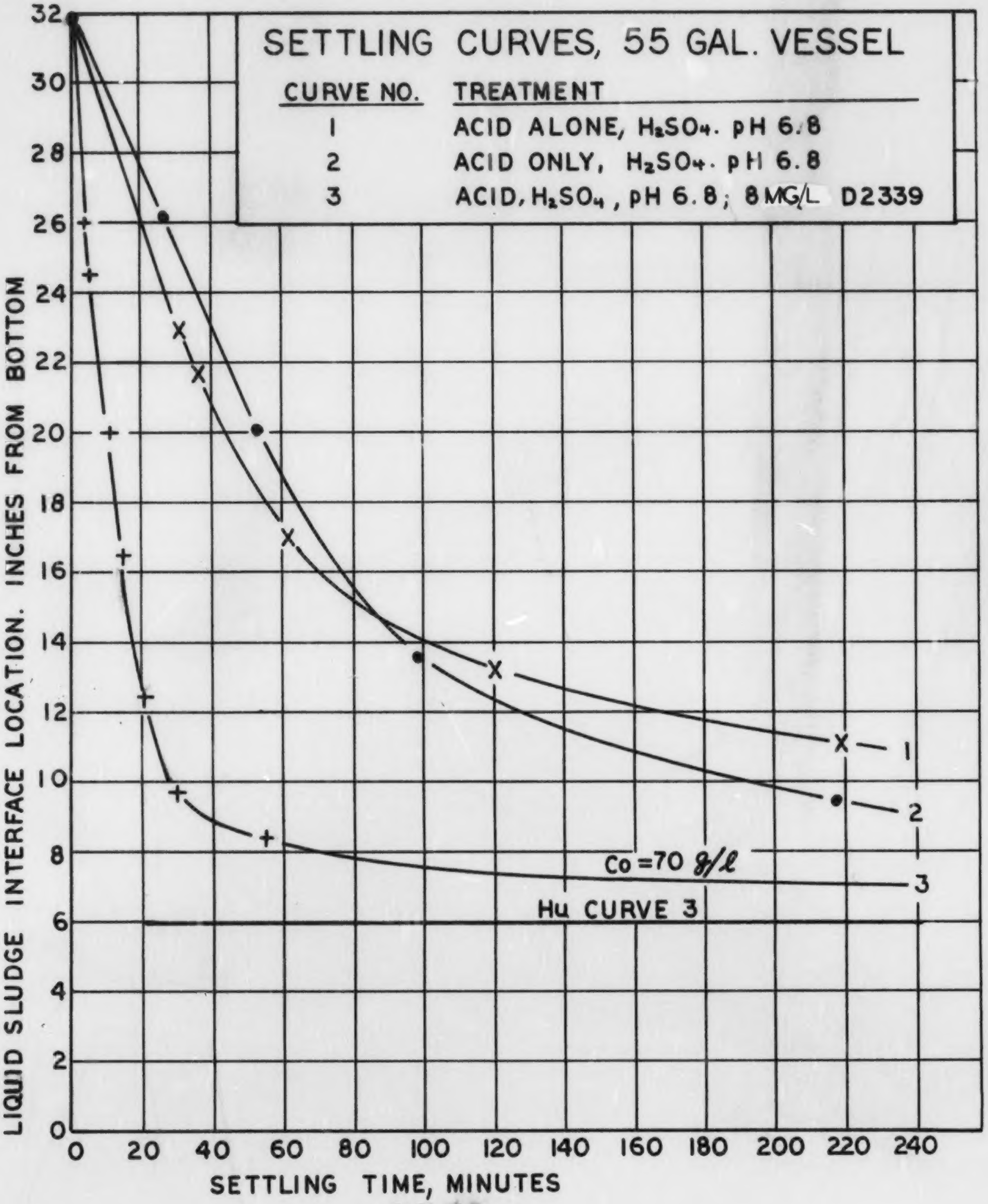
The settling tests used in calculation of thickener size were run in 55 gallon drums at the job site, i.e. next to sampling point C-8 (total flow from washery). The barrels were filled with waste from the coal washery effluent. The pH was adjusted immediately with sulfuric acid pickle liquor, (60,000 mg/l mineral acidity), while the barrel was agitated manually. After approximately 3-5 minutes of agitation, D-2339, a high molecular weight anionic polyelectrolyte, was added. Gentle agitation was continued for another 3-5 minutes while a sample of the treated suspension was collected from the barrel. The barrel was then allowed to stand undisturbed while settling occurred. The results are found in Table 54, Page 176.

Location of the sludge-liquid interface was determined by siphoning liquid from the barrel through a glass tube approximately 3/16" diameter. The tube was lowered slowly into the barrel in order to locate the interface. As soon as sludge appeared in the glass tube, siphoning was discontinued. The distance from the top of the barrel to the tip of the siphon was measured and then subtracted from the height of the barrel to determine the depth of the interface. Direct observation of the interface was impossible because fine coal collected at the surface of the liquid.

The barrels were allowed to stand for as much as 36 hours in order to determine final sludge level. Samples of the supernate were collected from each barrel at the end of the test.

Settling data are shown in Figures 57-59. The curves were used for calculation of thickener unit area, ft^2/min-lb. Overflow rate, in terms of gallons per minute per square foot, was then calculated from the unit area. The method of calculation of unit area was made in accordance with Standard Procedures as outlined by Talmage and Fitch[2]. Overflow rates were calculated on the basis of material balance of water and solids through the thickener. Results of these calculations are shown in Table 56.

The data shown on Figures 58 and 59 clearly indicate that a polyelectrolyte such as Nalco D-2339 will be required if the treated waste is to be handled economically in a clarifier. As shown in Table 56, the overflow rates even with the use of this polyelectrolyte are also very low. Overflow rates below 0.2 gpm/ft^2 are considered to be low. Omitting the polyelectrolyte drastically reduces the settling rates and makes the use of a clarifier completely impractical.



SETTLING CURVES, 55 GAL. VESSEL
CURVE NO.
TREATMENT
1
ACID ALONE, H₂SO₄. pH 6.8
2
ACID ONLY, H₂SO₄. pH 6.8
3
ACID, H₂SO₄, pH 6.8; 8 MG/L D2339
Co = 70 g/l
Hu CURVE 3
1
2
3
32
30
28
26
24
22
20
18
16
14
12
10
8
6
4
2
0
0
20
40
60
80
100
120
140
160
180
200
220
240
LIQUID SLUDGE INTERFACE LOCATION. INCHES FROM BOTTOM
SETTLING TIME, MINUTES

FIG. 57

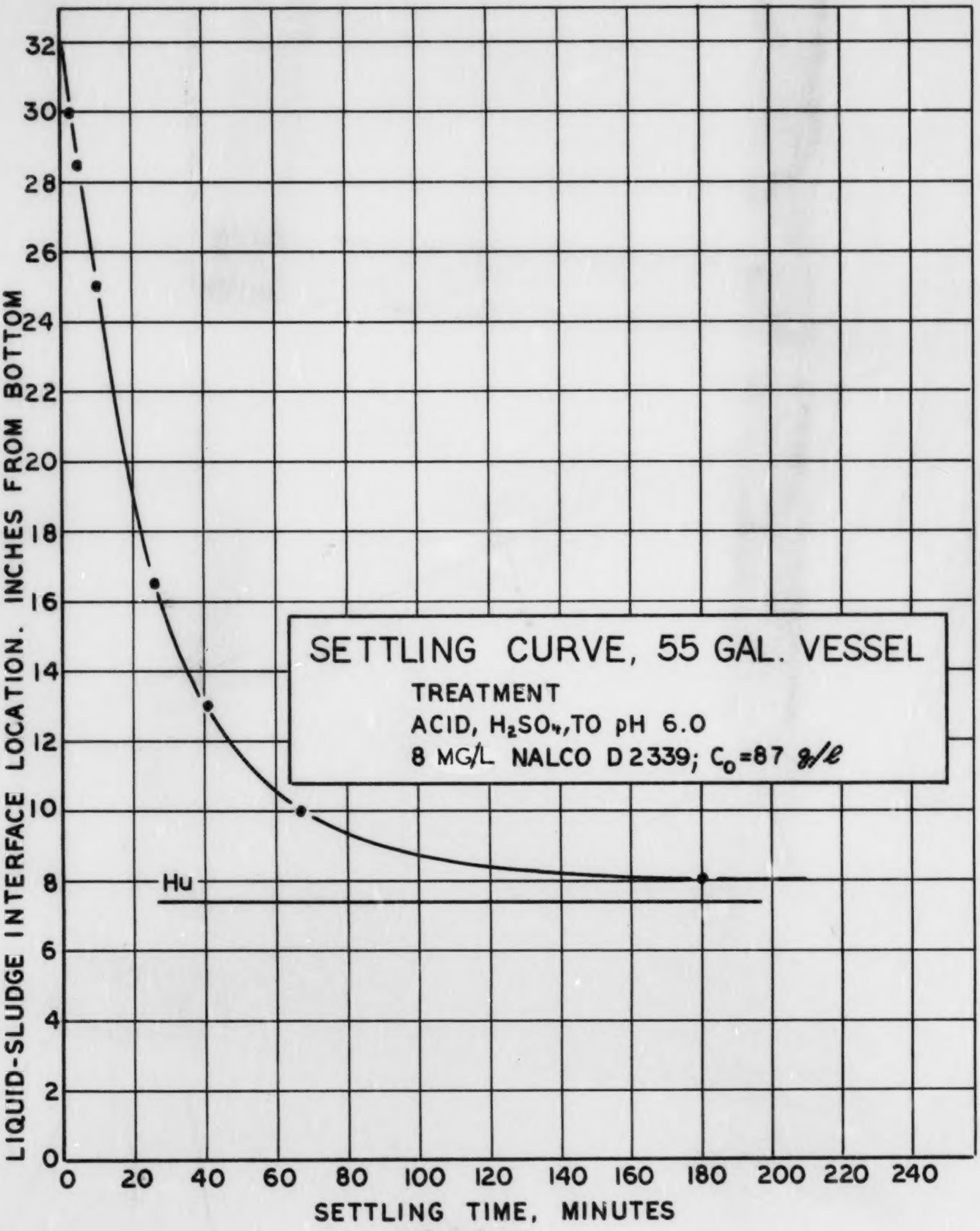

SETTLING CURVE, 55 GAL. VESSEL
TREATMENT
ACID, H_2SO_4, TO pH 6.0
8 MG/L NALCO D2339; C_0=87 g/l
Hu
LIQUID-SLUDGE INTERFACE LOCATION. INCHES FROM BOTTOM
0
2
4
6
8
10
12
14
16
18
20
22
24
26
28
30
32
0
20
40
60
80
100
120
140
160
180
200
220
240
SETTLING TIME, MINUTES

FIG 58

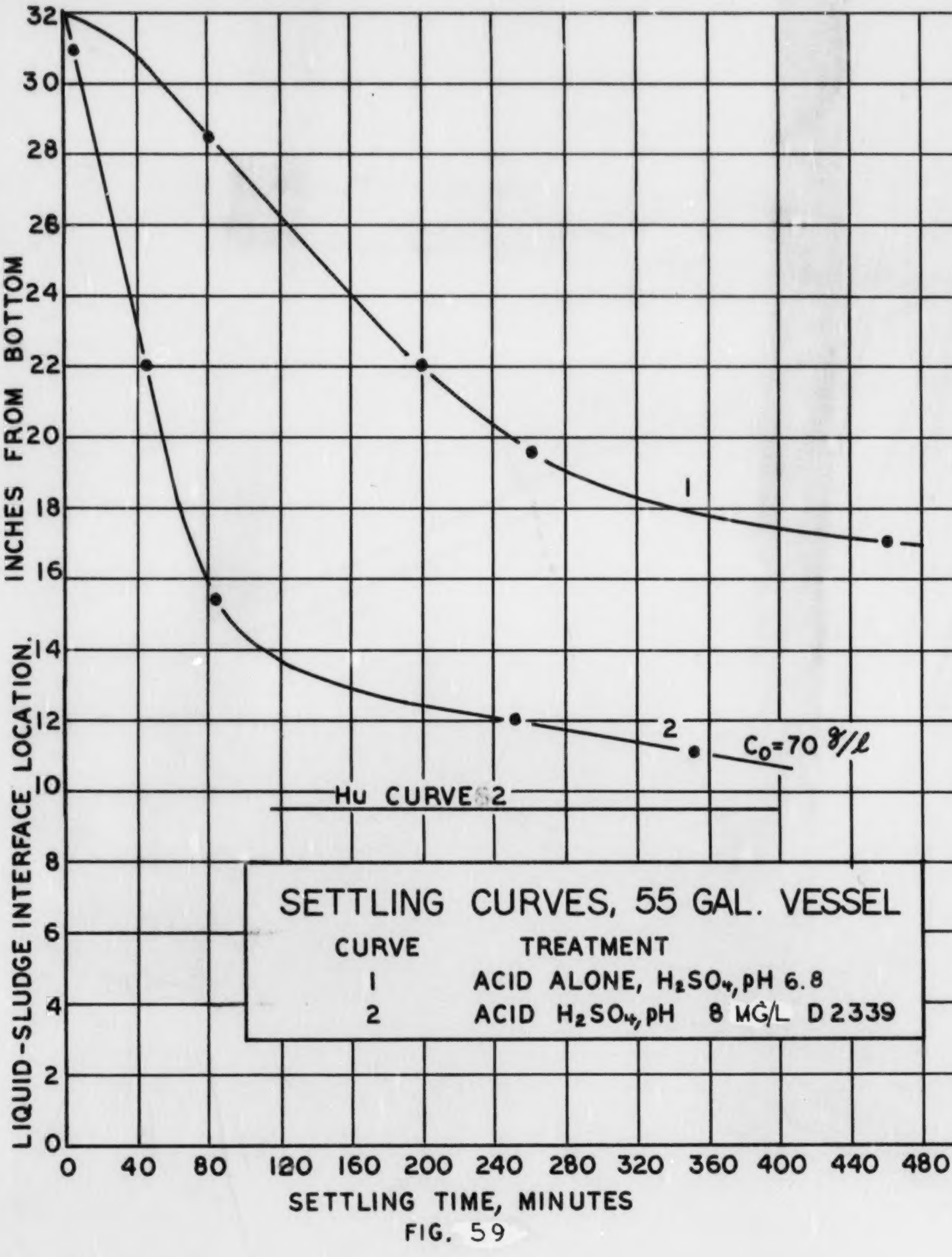

SETTLING CURVES, 55 GAL. VESSEL
CURVE TREATMENT
1 ACID ALONE, H_2SO_4, pH 6.8
2 ACID H_2SO_4, pH 8 MG/L D2339
Hu CURVE 2
C_0=70 g/l
1
2
LIQUID-SLUDGE INTERFACE LOCATION. INCHES FROM BOTTOM
32
30
28
26
24
22
20
18
16
14
12
10
8
6
4
2
0
0
40
80
120
160
200
240
280
320
360
400
440
480
SETTLING TIME, MINUTES

FIG. 59

TABLE 56

CALCULATION OF OVERFLOW RATES FOR COAL WASHERY CLARIFIER

A. Settling Test on August 5.
See settling curve on Figure 57, Curve 3,

Suspended solids concentration prior to settling, $C_o = 70$ g/l
Initial suspension height, $H_o = 32$ inches
Ultimate height of sludge, $H_u = 6$ inches

Since $C_oH_o = C_uH_u$

$$C_u = \frac{32}{6} \text{ X } 70 = 353 \text{ g/l}$$

Calculation of unit area, $\frac{\text{ft.}^2 - \text{min.}}{\text{lb.}}$ (U.A.)

Compaction Rate $= t_u$

$$\text{U.S.} = \frac{t_u}{C_oH_o}$$

From Figure 57, $t_u = 100$ min.

$$\text{U.A.} = \frac{100 \text{ X } 12}{32} \text{ X } \frac{10^6}{70{,}000} \text{ X } \frac{1}{62.4} = 8.8 \frac{\text{ft.}^2 - \text{min.}}{\text{lb.}}$$

Calculation of overflow rate (O.R.)

$$\text{O.R.} = \frac{120}{\text{U.A. x } C_o} \text{ X } \frac{C_u - C_o}{C_u}$$

$$\text{O.R.} = \frac{120}{8.8 \text{ X } 70} \text{ X } \frac{353 - 70}{353} = 0.157 \text{ gpm/ft.}^2$$

B. Settling test on November 20.
See Figure 58.

$C_o = 87$ grams/liter
$H_o = 32$ inches
$H_u = 7.5$ inches
$T_u = 100$ min.

$$C_u = \frac{32}{7.5} \text{ X } 87 = 372 \text{ g/l}$$

$$\text{U.A.} = \frac{100 \text{ x } 12}{32} \text{ x } \frac{10^6}{87{,}000} \text{ x } \frac{1}{62.4} = 6.95 \frac{\text{ft.}^2 - \text{min.}}{\text{lb.}}$$

$$\text{O.R.} = \frac{120}{6.95 \text{ x } 87} \text{ x } \frac{372 - 87}{372} = .153 \frac{\text{gpm}}{\text{ft.}^2}$$

As seen in Table 56, two settling tests gave a thickener overflow rate of approximately .15 gallons per minute per square foot. It would appear, therefore, that the design overflow rate of the thickener required to handle the washery effluent should be in the neighborhood of .15 gpm per square foot, static conditions. At a flow rate of 600 gpm, this is equivalent to a clarifier size of approximately 5,000 square feet, or 80 feet effective diameter.

The total sludge volume resulting from clarifier operation will amount to levels between 24% and 38% by volume. Alternatives for sludge handling are vacuum filtration, centrifugation, or impounding followed by gravity settling and supernate decantation.

Dewatering by Vacuum Filtration

Tests were run in a continuous system using an Eimco belt type vacuum filter three feet in diameter by one foot in face length. The system was erected in the coal washery and operated directly off the discharge from the tailings tank. (See Figure 60).

A pilot size air flotation unit was used for separation of fine coal from the tailings since it was felt that the presence of coal particles might interfere with the performance of the vacuum filter. The effluent from the flotation unit overflowed to an agitation tank where acid was added by gravity from an overhead reservoir. The pH control was manual. The resulting effluent was run to a second tank where the polyelectrolyte, Nalco D-2339, was added with slow agitation. The suspension overflowed to a settling tank for solids-liquid separation. A continuous overflow from the settling tank was sent directly to the sewer while the solids accumulated to a volume sufficient to begin pumping to the vacuum filter. After solids accumulation, the supernate was decanted. Results of the pilot vacuum filter are shown in Table 57.

The performance of the vacuum pilot filter was not satisfactory because of poor solids removal obtained per unit belt area, thin cake layers on the belt, and high lime costs. A 30% by weight addition of lime or calcium carbonate was necessary in the pilot system in order to obtain adequate cake release. Operation of the system without lime feed to the slurry resulted in extremely wet cakes which gave very poor or no release from the filter belt. These results were confirmed through vacuum filter leaf tests conducted on coal washery samples sent to the Eimco Corporation.

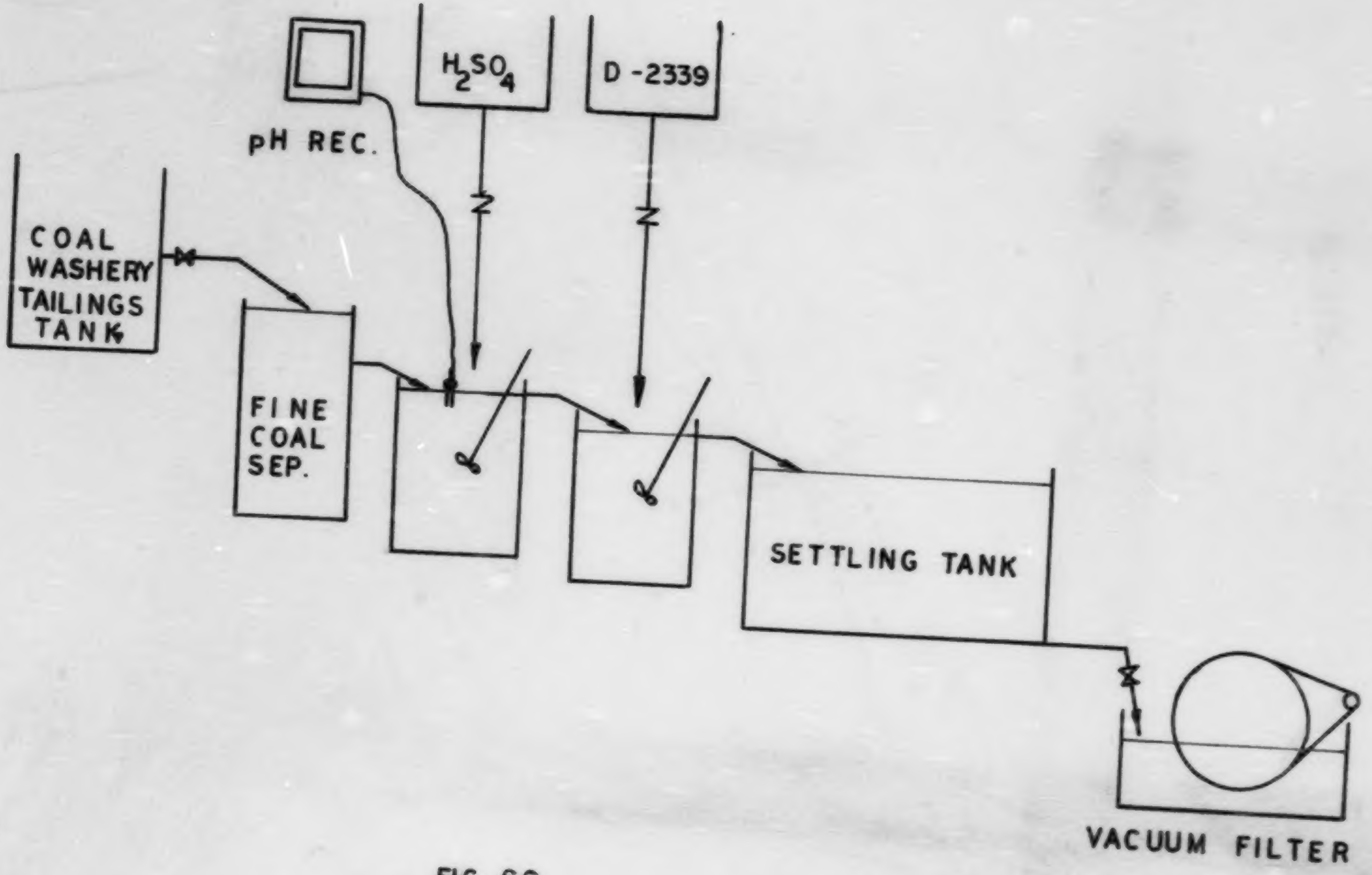

pH REC.
H_2SO_4
D -2339
COAL WASHERY TAILINGS TANK
FINE COAL SEP.
SETTLING TANK
VACUUM FILTER

FIG. 60
PILOT UNITS. COAL WASHERY EFFLUENT TREATMENT SYSTEM

TABLE 57

RESULTS OF PILOT UNIT VACUUM FILTER TESTS

Date	Pretreatment Pct. by wt.	Solids Basis	Run No.	Vacuum In. Left	Hg. Rt.	Cake Thickness Inches	Drum Speed Sec/Ft	Cake Weights Wet Grams	Dry Grams	% Moisture	% Ash in Solids	Cake Release
Feed Concentration 35% Solids												
2/5/70	-	-	0	19	20	1/8	30	75.5	53.7	28.9	63.1	-
	-	-	1	19	20	1/8	30	72.0	53.5	25.7	64.5	-
	-	-	2	19	20	1/8	26	68.0	48.5	28.7	65.4	-
	-	-	3	19	20	1/8	26	62.2	43.9	29.4	63.1	-
	-	-	4	19	20	3/32	30	59.1	43.0	27.2	63.6	
	-	-	5	-	-	-	30	47.8	35.5	28	63.6	-
	-	-	6	-	-	-	-	48.3	35.6	28.0	67.1	-
	Hyd. Lime	(unmeasured	7	-	-	-		126.4	89.5	29.1	66	Good
	Hyd. Lime	quantity)	8	-	-	-	-	122.4	89.0	27.3	65.5	Good
Feed Concdntration 28% Solids												
2/13/70	5% $CaCO_3$	-	1	21	19	1/16	30	40.2	31.0	25.1	73.0	Poor
	5% $CaCO_3$	-	2	21	19	1/16	30	34.9	26.9	25.9	72.8	Poor
		10% $CaCO_3$	3	20	18	1/8	30	58.1	45.0	25.9	74.4	Good
		20% $CaCO_3$	4	20	18	1/8-3/16	30	65.4	51.4	22.5	73.7	Good
		4 lbs. Diat. Earth	5	20	18	1/8-1/4	30	62.7	46.1	28.1	71.8	Good
	5% $CaCO_3$	-	6	20	18	1/16-1/8	38	53.9	36.3	34.9	72.6	Poor
		10% CaO	7	20	18	1/16-1/8	36	33.0	23.1	33.3	73.1	Good
		10% CaO	7A	20	18	1/8	26	50.2	34.9	31.9	72.9	Good
		20% CaO	8	20	18	1/8	30	57.3	40.4	32.1	73.5	Good
		10 lbs. Diat. Earth	8A	20	18	1/8	30	65.5	47.6	29.5	75.0	Good
Feed Concentration 17% Solids												
2/17/70	5% Hyd. Lime	-	1	19	21	1/16	28	46.2	29.9	37.0	74.6	Poor
	10% Hyd. Lime	-	2	19	21	1/16 1/8	26	51.9	32.4	39.7	74.8	Poor
	18% Hyd. Lime		3	18	20	1/8	26	49.7	31.6	38.3	74.8	Poor
	18% Hyd. Lime		4	18	20	1/8	28	58.1	43.1	38.1	74.5	Poor
	30% Hyd. Lime	-	5	18	20	1/8	22	52.7	34.2	37.1	74.7	Poor
Feed Concentration 17% Solids												
2/20/70		-	1	18	20	1/16-1/8	30	(Too Wet for Sampling)				
	11.1% CaO	-	2	18	20	1/16-1/8	30	(Too Wet for Sampling)				
	29.4% CaO	-	3	18	30	1/16-1/8	30	36.7	26.6	30.6	74	Poor
		-	4	18	30	1/16-1/8	30	49.6	35.6	30.4	73.2	Poor
		-	5	18	30	1/8	30	58.6	44.5	25.5	74.1	Good
		-	6	18	30	1/8-3/16	30	74.9	56.4	25.8	75.3	Good
		-	7	18	30	3/16-1/4	30	97.8			75.5	Good
		-	8	18	30	1/4	30	128.9	-	25.2	75.4	Good
		-	9	18	30	1/4	30	99.6	-	-	76.4	Good
		-	9A	18	30	1/4	26	-	-	-	-	Good
		-	9B	18	30	3/16	15	-	-	-	-	Good
		-	9C	18	30	1/8	9	-	-	-	-	Poor
		-	10	18	30	1/4	30	183.8	140.3	24.0	80.0	Good
Heavy Sludge		-	1	-	-	-	-	160.9	140.2	12.2	80.5	-
Heavy Sludge		-	2	-	-	-		105.8	73.3	22.3	81.0	-

Since the vacuum filtration tests indicated unsatisfactory and expensive operation of such a device, its use for the final system was discounted and other disposal methods were considered for handling the clarifier underflow. Screening tests with a laboratory centrifuge operated at low speeds gave excellent dewatering and a clear supernate. The centrifuge method was later determined as the most desirable for sludge dewatering. Further discussion and evaluation of the centrifuge and inponding methods of sludge disposal are contained in the Solid Waste Investigation section, pages 211-238.

PROCESS MODIFICATION

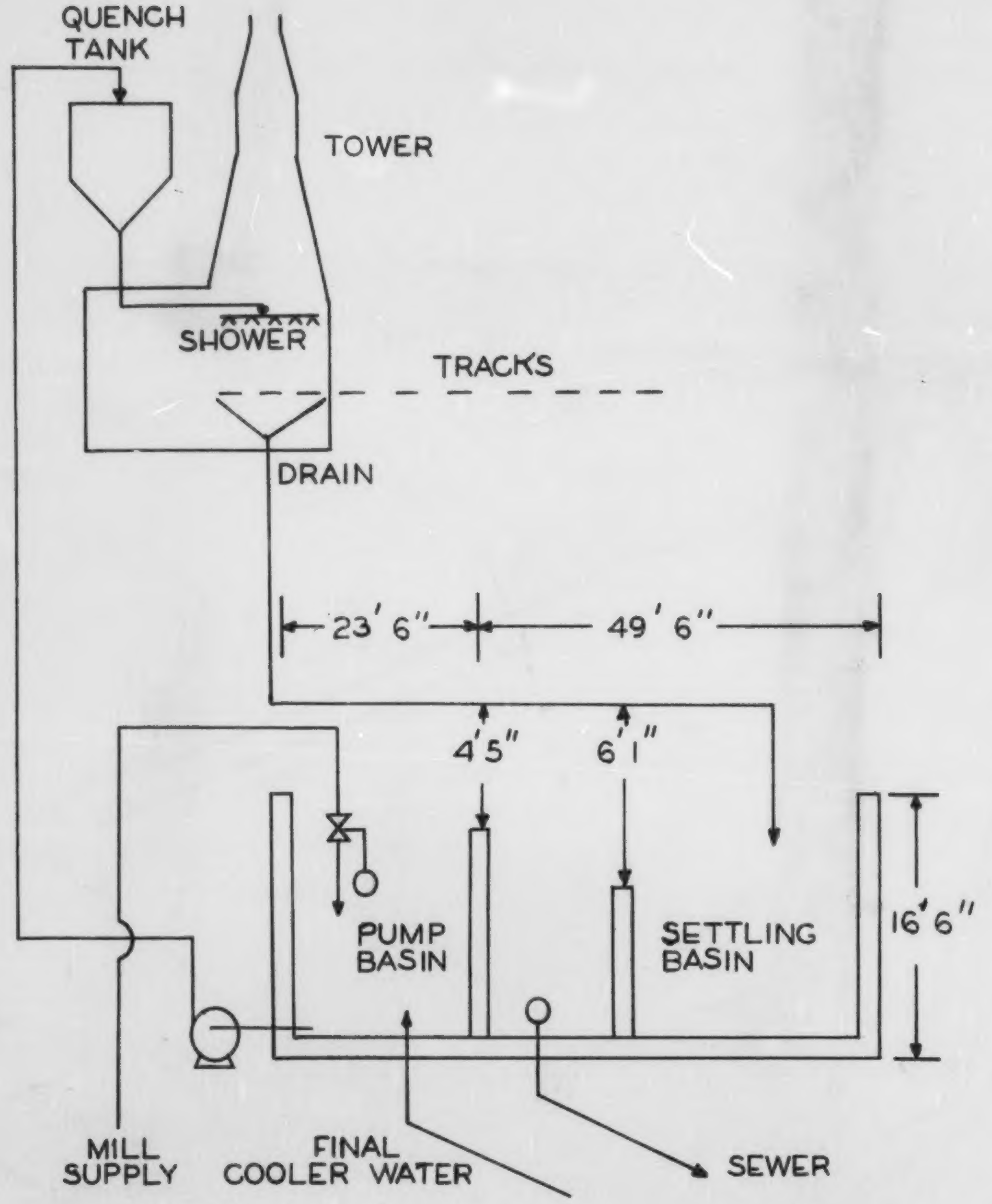
Study of Quench Station Operation

The plant has two quench stations where the incandescent coke pushed from the ovens is cooled by means of water sprays. Because of their location within the plant these stations are designated North and South quench stations. Figures 61 and 62 show the schematic layout of these stations. These stations have traditionally operated on a once through water flow basis. Used water from the final cooler in the by-product plant is roughly split into two equal flows by manual valve adjustment with one half being supplied to each quench station. In addition, regular plant makeup water is supplied to the pump basin of each station for makeup. This makeup water is regulated by a float controlled valve at the North station whereas the South station makeup is constant by means of a manually controlled valve. After use in the quench, the water not consumed drains to the settling basin for removal of the large concentration of coke breeze. The overflow passes to the plant sewer system and ultimately to point C-7A. Because of the water system, water overflows to the sewer constantly at the South station and intermittently at the North. However, since the North station is used more frequently, the total overflow normally averages 500 gpm from the North station and 300 gpm from the South. The spent quench water is contaminated with ammonia, phenol, and cyanide.

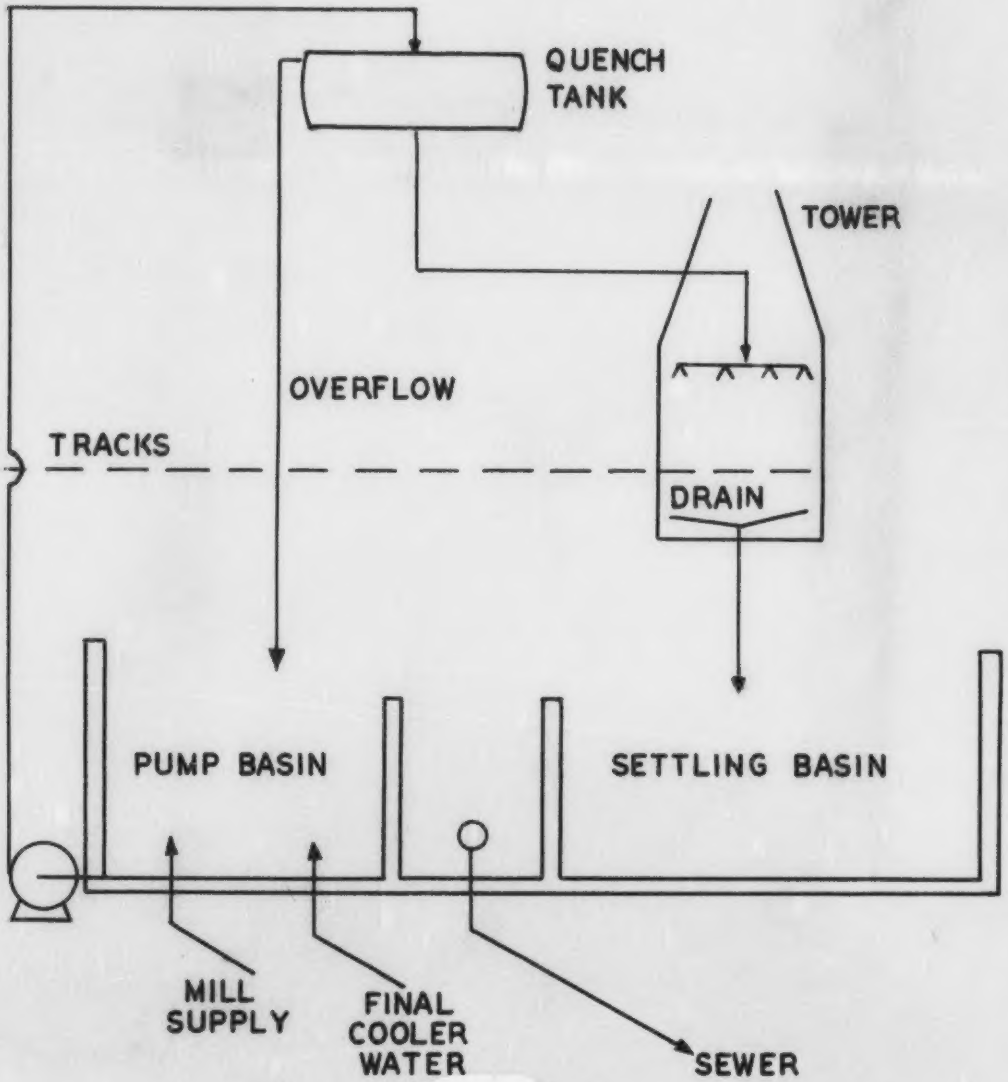
Reduction in water consumption at the two quenching stations would result in a corresponding reduction of both total pollution load and water volume at the combined plant discharge point (C7A). Such reduction could be accomplished by re-use of the spent quench water either completely or partly, as needed, in order to avoid scaling or corrosion problems common to the quenching facilities.

The quantity of water used for quenching at the South station was determined directly by measuring the drop of liquid level in the overhead quench tank, starting with the tank full and with the quench water supply pump turned off. Except for the volume determination, no additional testing was done at the South quench because of the less frequent use of this station and also because of the high rates of fresh mill water overflow.

The North quench station was not suited for determination of quench volume in the overhead tank since the tank was not fitted with an access opening. Instead the quench volume was


QUENCH TANK
TOWER
SHOWER
TRACKS
DRAIN
23' 6"
49' 6"
4' 5"
6' 1"
16' 6"
PUMP BASIN
SETTLING BASIN
MILL SUPPLY
FINAL COOLER WATER
SEWER

NORTH QUENCH STATION
FIG. 61


QUENCH TANK
TOWER
OVERFLOW
TRACKS
DRAIN
PUMP BASIN
SETTLING BASIN
MILL SUPPLY
FINAL COOLER WATER
SEWER

FLOW DIAGRAM
SOUTH QUENCH STATION

FIG. 62

determined by measurements in the pump basin. The liquid level was established before and after a quench and after the overhead tank had been refilled by the supply pump. The volume of water used per quench was determined by measuring the liquid level rise after the pump was shut off, the quenches in that period, the liquid level drop during the pumping period, and the pumping time.

The evaporation and wetting loss per quench was determined by allowing the quench to operate as a closed system; that is, the water supply and sewer drain were temporarily turned off. Under these conditions the water from the quench tower was returned to the supply sump for re-use. Since no water was coming into the sump from the outside, the liquid level in the sump decreased proportional to the evaporation from the coke wetting. Total loss was calculated from the drop in liquid level and the number of quenches handled by the system. Make up water was added periodically to refill the sump without overflow.

The changes in water quality, which resulted from the operation of the quenches as a closed system, were observed and recorded. Parameters included pH, chloride, hardness, temperature, conductivity. Corator readings (See Appendix, Anal. Methods) described the changes in corrosion or scaling tendencies. Additional parameters included potentiometric alkalinity, cyanide, and ammonia.

Table 58 shows a tabulation of all quench water volumes as determined for the two stations. As much as 9,000 gallons per quench was used at the North station. Variation in volume is due to operation. The South station appears to be using less water. This station is used in quenching coke from ovens of two sizes which explains some of the variation in volumes.

Table 59 shows results of analyses of water samples taken from the North quench station on December 4, 1969, when the station was operated on a once through basis with no recirculation of spent quench water. These results provide a basis for comparison with the closed system operation (recirculation).

Results for a closed recycle system are shown on Tables 59-63. The tables show the time of sampling, the number of quenches handled by the system during the time intervals indicated, and the use of final cooler water. Data on the recirculation system shows that the North quench station handled 6 quenches between ten and eleven the morning of December 4. The final cooler water was off during this time. At eleven a.m. the final cooler was turned on to refill the basin and left running until 2 p.m. A total of 12 quenches were handled between 2 and 4 p.m. The total liquid drop in the basin was measured for this period and the volume loss, gallons per quench, was calculated by the method shown in Table 58. At this time the system was changed from a closed recirculation to the normal once-through type operation.

Tables 62-63 show the data obtained during a 24 hour period of recirculation. The system was actually left running on a recirculation basis until December 11, at 8:30 a.m. At this time the system was returned to normal once through operation.

Figures 63 and 64 show titration curves for samples taken at the times indicated for the purpose of obtaining total alkalinities. The values for alkalinities were used in calculation of the Langelier Index shown in the tables.

The data reveals that 2,000-2,300 gallons of water is lost from the system with each quench. The loss is by way of steam going to the atmosphere and by the water which adheres to the coke after quenching. No analyses of the steam or coke moisture were conducted.

The data reveals no accumulation of total dissolved solids or any of the pollutants such as ammonia, cyanide, and phenols with continued operation of the quenching station as a closed system.

There is a definite trend of increasing reduction of alkalinity as the number of quenches increases. This is the main factor in reducing the scaling tendencies of the water as indicated by lower positive values of the Langelier Index. The Langelier Index of the quench water in the once through system as shown in Table 59 was 2.06. Langelier values during the period of recirculation were 1.41-1.77 (Tables 59-62). Maintenance records for the quench station indicate that the water is slightly corrosive under normal conditions of operation. The fact that the analytical data indicates a reduction in positive values of the Langelier Index means that the water is slightly more corrosive after the system has been closed up.

The Corator values shown in the tables confirm that higher rates of corrosion coincide with closing the system. The average rate of corrosion in the once-through system was 1.1 mils/year while an average corrosion rate in the closed recirculating system was 1.4 mils/year. (See Tables 59-63). However, both the Langelier Index as calculated and the Corator readings are relative values and can hardly be transformed to practical terms such as replacement requirements for the quench car or the structural steel facilities in the system.

TABLE 58

Results of Quench Volume Measurements

1. Quench Volumes, Gal./Quench

Date	North Quench	South Quench
2-20-69	5,980	5,500
2-20-69	-	5,300
6-02-69	9,000	3,630
6-02-69	-	4,520
7-14-69	8,050	5,600
7-14-69	8,300	-

2. Losses, Gals./Quench

Date		
12-4-69	2,280	-
1-15-70	2,000	-

TABLE 59

OPERATION			ANALYSES ON QUENCH WATER FOR 12-4-69 (NORMAL ONCE THROUGH OPERATION)									
Time of Interval	Number of Quenches	Cooler Water	pH	Temp °F	Conductivity in Micromho CM	Chloride as NaCl	Total Hardness as $CaCO_3$	Ca Hardness as $CaCO_3$	Corrosion in Mils/Year	Cyanide as CN	Ammonia as NH_3	Phenol as C_6H_5OH
1:43P 2:05P	3	On										
2:00P	-	-	8.8	152	1200	570	100	42	0.5	24	250	16
3:00P	-	-	8.8	162	1100	600	98	44	0.8	20	250	16
3:08P 4:04P	5	On										
4:00P	-	-	8.9	162	1000	550	96	42	2.1	21	250	20
4:04P 5:12P	6	On										
5:00P*	-	-	8.8	162	950	480	116	30	1.2	19	200	14
* Langlier Index + 2.06												
			ANALYSES ON QUENCH WATER FOR 12 2 69 (CLOSED OPERATION)									
1:00P 2:00P	5	On										
2:00P	-	-	8.5	142	1500	450	120	38	0.6		150	
* Langlier Index + 1.41												
			ANALYSES ON QUENCH WATER FOR 12-4-69 (CLOSED OPERATION)									
10:00A* 11:00A	6	Off	8.8	136	1900	-	84	26	0.8	22	250	15
11:00A	-	-	8.8	140	1800	-	80	26	0.3	-	200	-
11:00A 2:00P	17	On										
2:00P	-	-	8.8	151	1100	-	82	26	1.8	14	200	15
2:00P 3:00P	4	Off										
3:00P	-	-	8.8	154	950	-	82	32	1.5		150	1
3:00P 4:00P	6	Off										
4:00P**	-	-	8.7	154	975		100	40	1.6	5.0	100	11

* Langlier Index + 2.1
** Langlier Index + 1.51

TABLE 60

OPERATION

ANALYSES ON QUENCH WATER FOR 12-8-69 (CLOSED OPERATION)

Time of Interval	Number of Quenches	Cooler Water	pH	Temp °F	Conductivity in Micromho CM	Chloride as NaCl	Total Hardness as $CaCO_3$	Ca Hardness as $CaCO_3$	Corrosion in Mils/Year	Cyanide as CN	Ammonia as NH_3	Phenol as C_6H_5OH
2:15P* 3:15P	8	Off	8.9	154	920	480	86	26	4.9	15	150	23
3:15P	-	-	8.9	154	920	440	88	32	3.1	-	150	
3:15P 5:00P	8	Off										
5:00P	-	-	8.2	154	800	550	86	26	4.6	14	200	20
5:00P 8:00P	7	Off										
8:00P 12:00A	17	On										

* Langlier Index +1.74

ANALYSES ON QUENCH WATER FOR 12-9-69 (CLOSED OPERATION)

Time of Interval	Number of Quenches	Cooler Water	pH	Temp °F	Conductivity in Micromho CM	Chloride as NaCl	Total Hardness as $CaCO_3$	Ca Hardness as $CaCO_3$	Corrosion in Mils/Year	Cyanide as CN	Ammonia as NH_3	Phenol as C_6H_5OH
12:00A 8:30A	38	On										
8:30A	-	-	8.8	150	1010	400	92	26	1.6	30	250	19
8:30A 10:00A	3	On										
10:00A	-	Off	9.0	143	1200	600	92	28	1.4	-	250	-
10:00A 11:00A	6	Off										
11:00A	-	-	8.9	144	1490	540	102	30	1.3	16	200	24
11:00A 1:00P	10	Off										
1:00P*	-	On	8.6	142	1500	430	136	44	0.8	7.0	100	25
2:00P 3:00P	6	On										
3:00P	-	-	9.0	140	1500	480	108	36	0.8	-	200	-
3:00P 4:30P	7	On										
4:30P	-	-	8.8	148	1490	510	110	36	0.8		200	-
4:30P 5:30P	6	On										
5:30P	-	-	8.8	148	1490	560	110	30	1.1	22	250	32
5:30P	10	On										

TABLE 61

OPERATION			ANALYSES ON QUENCH WATER FOR 12-9-69 (CLOSED OPERATION) (CONT.)									
Time of Interval	Number of Quenches	Cooler Water	pH	Temp ·F	Conductivity in Micromho CN	Chloride as NaCl	Total Hardness as $CaCO_3$	Ca Hardness as $CaCO_3$	Corrosion in Mils/Year	Cyanide as CN	Ammonia as NH_3	Phenol as C_6H_5OH
7:30P												
7:30P	-	Off										
10:00P 12:00A	9	Off										
			ANALYSES ON QUENCH WATER FOR 12 10 69 (CLOSED OPERATION)									
12:00A 1:50A	10	On										
1:50A 4:20A	10	Off										
4:20A	-	On										
4:20A 6:15A	12	On										
6:15A	-	Off										
6:15A 8:30A	10	Off										
8:30A	-	-	9.0	162	800	360	116	40	1.1	8.0	150	30
8:30A 10:20A	9	Off										
3:00P	-	Off										
3:00P 4:00P	3	Off										
4:00P	-	-	9.2	154	1320	580	106	26	0.5		200	
4:00P 4:25P	5	Off										
4:25P	-	On										
5:00P	-	-	8.6	150	1300	460	142	42	0.6	6.0	150	29
5:00P 6:45P	10	On										
6:45P	-	Off										
6:45P 9:45P	14	Off										
12:00A	13	On										
10:20A 10:30A	1	On										

TABLE 62

OPERATION

ANALYSES ON QUENCH WATER FOR 12-10-69 (CLOSED OPERATION) (CONT.)

Time of Interval	Number of Quenches	Cooler Water	pH	Temp °F	Conductivity in Micromho CM	Chloride as NaCl	Total Hardness as $CaCO_3$	Hardness as $CaCO_3$	Corrosion in Mils/Year	Cyanide as CN	Ammonia as NH_3	Phenol as C_6H_5OH
10:30A*	-	-	8.9	160	775	360	162	54	0.8	-	100	-
10:30A 12:30P	12	On										
12:30P	-	-	8.6	166	775	420	130	34	0.5	13	150	27
12:30P 2:00P	2	On										
2:00P	-	-	8.9	152	1200	470	114	36	1.0	-	175	-
2:00P 3:00P	5	On										

* Langlier Index +1.77

ANALYSES ON QUENCH WATER FOR 12 11 69 (CLOSED OPERATION)

12:00A 1:30A	7	On										
1:30A	-	Off										
1:30A 3:45A	11	Off										
3:45A	-	On										
3:45A 5:20A	8	On										
5:20A	-	Off										
5:20A 7:56A	13	Off										
7:56A	-	On										
7:56A 8:30A	3	On										
8:30A		-	9.1	160	900	350	160	68	1.1	3.0	100	22
8:30A 10:00A	8	On										
10:00A	-	-	9.2	156	830	450	110	40	-		150	-
10:00A 12:00P	7	On										
12:00P	-	Off										
12:00P 1:00P	7	Off										
1:00P	-	-	9.2	154	1100	570	92	32	-	-	250	38

TABLE 63

ANALYSES ON QUENCH WATER FOR 12-11-69 (CLOSED OPERATION) (CONT.)

OPERATION												
Time of Interval	Number of Quenches	Cooler Water	pH	Temp F	Conductivity in Micromho CM	Chloride as NaCl	Total Hardness as $CaCO_3$	Ca Hardness as $CaCO_3$	Corrosion in Mils/Year	Cyanide as CN	Ammonia as NH_3	Phenol as C_6H_5OH
1:00P 2:30P	4	Off										
2:30P	-	-	9.2	152	1090	640	120	40	-	-	200	-
2:30P 3:20P	6	Off										
3:20P	-	On										
3:20P 4:00P	4	On										
4:00P	-	-	8.9	152	1050	520	122	40	-	-	150	-
4:00P 5:00P	5	On										
5:00P	-	-	8.7	152	1050	600	120	32	-	-	200	-
5:00P 7:30P	14	On										
7:30P	-	Off										
7:30P 10:00A	13	Off										
10:00A	-	On										
10:00A 1:50A	19	On										
1:50A	-	Off										
1:50A 4:20A	14	Off										
4:20A	-	On										
4:20A 6:15A	9	On										
6:15A	-	Off										
6:15A 8:30A	9	Off										
8:30A	-	-	9.2	150	1030	530	94	28	0.0	-	300	-

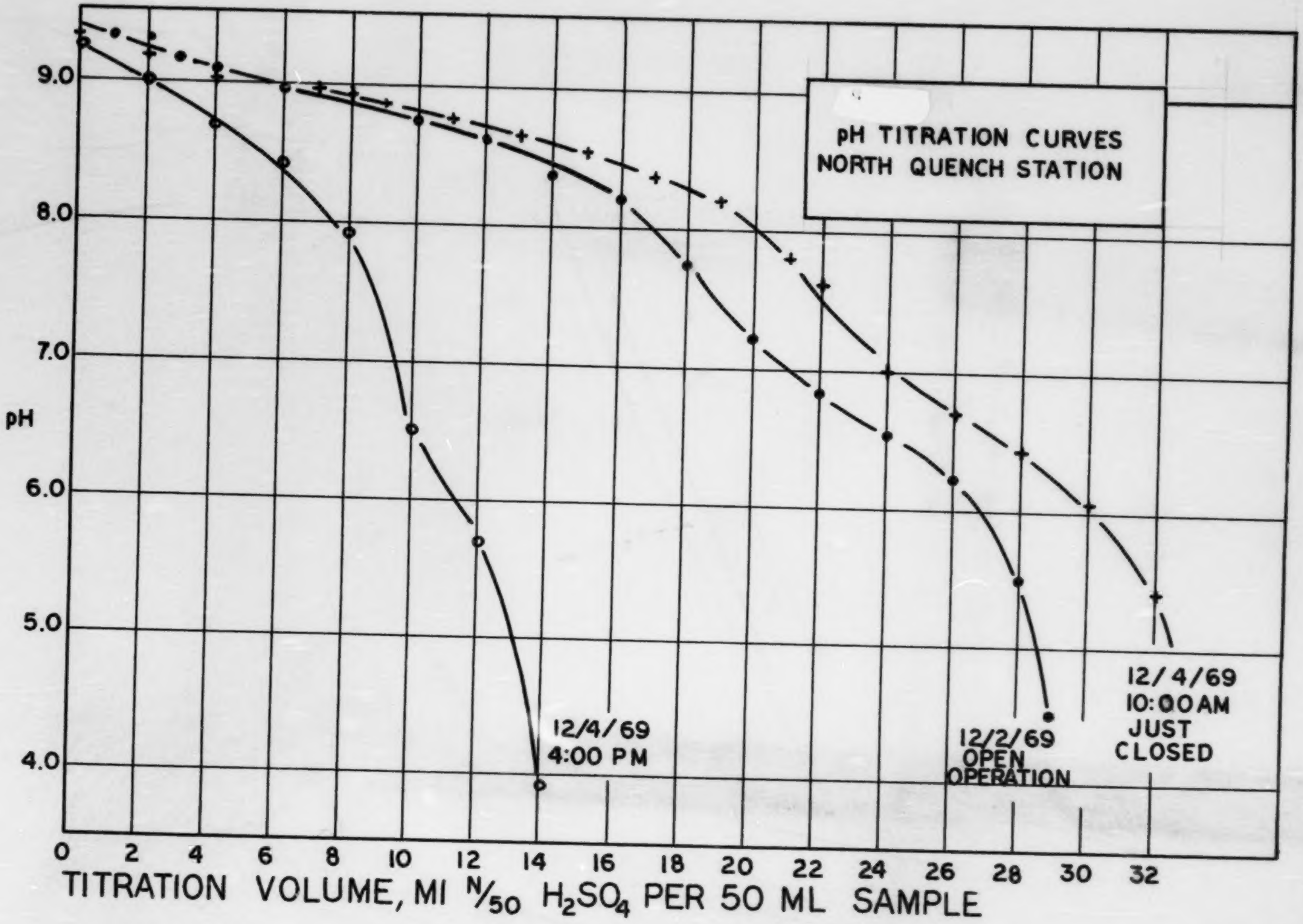

pH TITRATION CURVES
NORTH QUENCH STATION
pH
9.0
8.0
7.0
6.0
5.0
4.0
0
2
4
6
8
10
12
14
16
18
20
22
24
26
28
30
32
TITRATION VOLUME, Ml $^N/_{50}$ H_2SO_4 PER 50 ML SAMPLE
12/4/69
4:00 PM
12/2/69
OPEN
OPERATION
12/4/69
10:00AM
JUST
CLOSED

FIG. 63

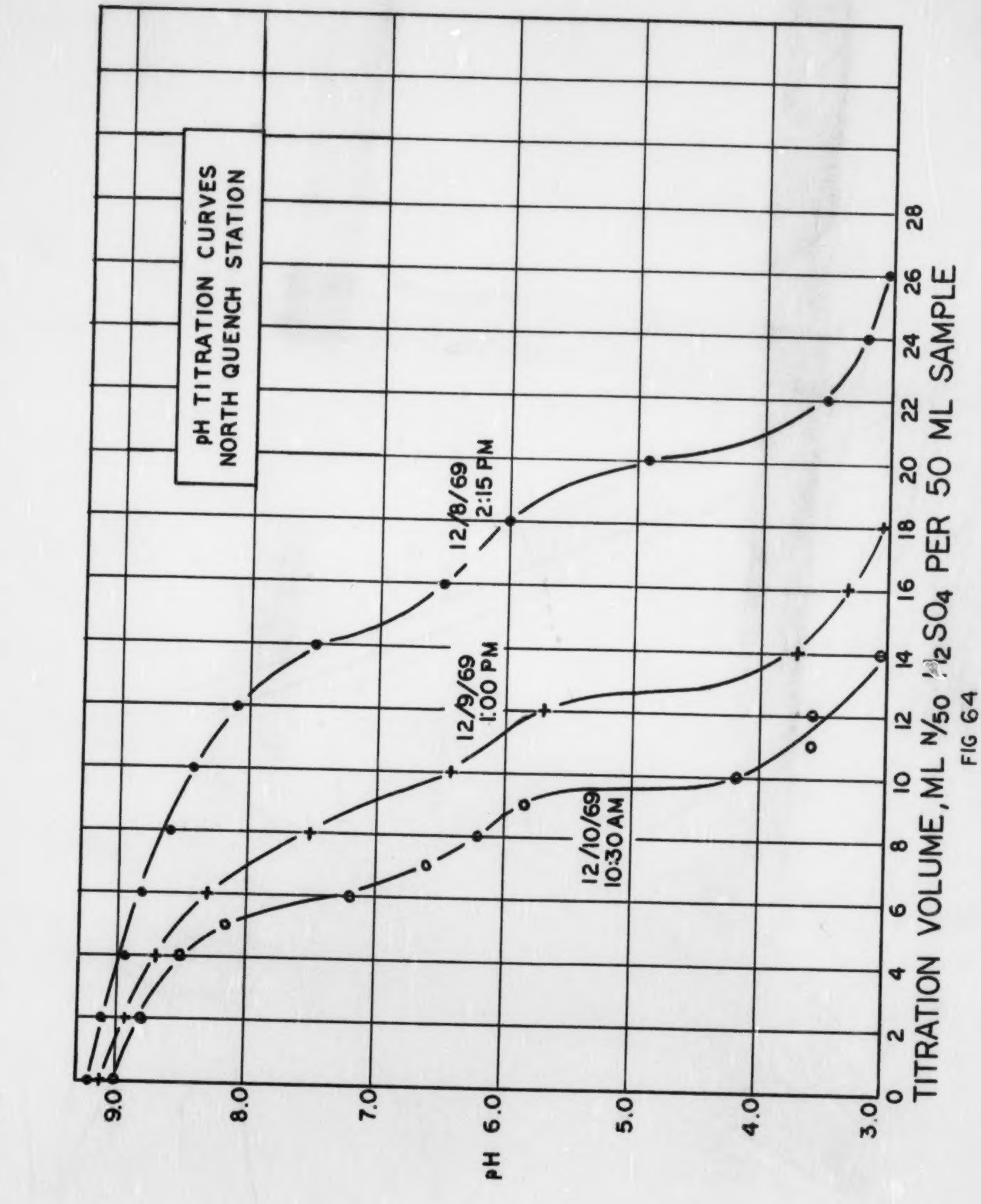

pH TITRATION CURVES
NORTH QUENCH STATION
12/8/69
2:15 PM
12/9/69
1:00 PM
12/10/69
10:30 AM
pH
9.0
8.0
7.0
6.0
5.0
4.0
3.0
0
2
4
6
8
10
12
14
16
18
20
22
24
26
28
TITRATION VOLUME, ML N/50 H_2SO_4 PER 50 ML SAMPLE

FIG 64

Corrosion has been only a very minor problem in the past. The very small changes in the corrosion index indicate no serious corrosion problem in the operation of the quenches on a closed system using either final cooler water or mill supply water makeup.

During normal coking operations 150-220 ovens per day are pushed and subsequently quenched. This represents a consumption of 300,000 and 500,000 gallons of water per day.

Caustic Dumps

Following the removal of sulfur compounds from secondary light oil by treatment with sulfuric acid in an agitator, the oil is neutralized with a solution of caustic soda. The caustic soda solution is mixed with water in a ratio of 7 parts water to 4 parts caustic soda prior to use. After the caustic solution is added to the light oil, agitation is continued for approximately 20 minutes to provide good mixing and neutralization. Agitation is then discontinued. The mixture is allowed to settle for another 25-30 minutes to allow for caustic and oil separation. The partially spent caustic solution is then wasted to the sewer. The washed light oil is removed for distillation into motor fuel or it is sent to the benzol fractionating column where crude benzol is produced. The spent caustic solution often contains large quantities of oil which deemulsify in the sewer. Additional undesirable effects caused by wasting caustic to the coke plant effluent are an increase in pH and the formation of yellowish, mild-white color when mixture of the streams occurs. Some of the oil collected in an interceptor sump in the southeast corner of the benzol yard, Figure 4, page 19, is emulsified by the waste caustic. The tests described below were run in order to determine the time needed for separation of all of the oil from the spent caustic solution.

One liter samples for the separation studies were obtained while the caustic waste was being discharged to the sewer. Samples were taken at the start of the dumping step and after several minutes of caustic discharge. The samples were identified by the time at which they were taken (i.e. minutes from starting of wasting).

The caustic oil samples were allowed to stand for as long as 19 hours. During this time the oil separated from the caustic layer and accumulated at the top of the samples. The volume of oil separated from the emulsion was measured and quantities were converted to percent of the total sample volume.

Test No. 1 showed a very gradual and prolonged process of separation. The benzol plant operator indicated that the batch was not normal, inasmuch as his product recovery and separation were very poor. The other two tests showed that ultimate separation was obtained in 1.5 hours. (See Table 64.)

The alkalinity remaining in the spent caustic solutions was negligible in comparison with the total amount of caustic used for neutralization. The volume of normal caustic solution (used for neutralization) required to raised 100 ml of acidified oil from pH 0.5 to pH 7.0 was 1.3 ml, while 54-62 ml. of the spent caustic was required to raise 100 ml. of acidified oil from pH 0.5 to pH 7.0.

Recycle of the waste caustic solution will not be considered from the standpoint of caustic savings, (neutralizing ability of the normal caustic being 50 times that of the waste caustic solution) but from the standpoint of economical handling of a waste stream. To prevent pollution of the coke plant waste stream by the spent caustic solution, the caustic will be discharged to a 300 gallon storage tank large enough to handle a single dump. As shown by the separation studies, 24 hours is sufficient time for the caustic-oil separation. The oil will be reclaimed by decanting and used to mix the next batch of caustic solution. Any sludge collected can be neutralized and disposed of as a solid.

Reuse of the spent caustic solution will require the use of simple titration tests by the operating personnel to properly adjust the caustic solution.

Coke Plant Ammonia Still Operation

The still consists of three principal parts through which the liquor flows: the "free" leg, the lime reservoir, and the "fixed" leg. In the free leg, ammonia nitrogen is rendered insoluble by the application of steam. Ammonia is volatilized together with hydrogen sulfide, carbon dioxide, hydrogen cyanide and a portion of the organic matter. The fixed and free legs of the still are made up of standard, cast iron, bubble-plate sections. The liquor is pumped continuously into the top of the free leg and flows downward from plate to plate. The liquor stream contacts the ascending current of steam. The latter is admitted at the base of the fixed leg, ascends through it and the free leg, but does not pass through the lime reservoir. The lime reservoir is a vertical, cylindrical chamber with a conical bottom and is composed of steel or cast iron. The lime and ammonium salts react here. A small quantity of steam is usually introduced at the base of the lime

TABLE 64

Results of Oil-Water Separation Tests on Caustic Dumps

Oil fractions, reported in percent

Test No. 1 Feb. 18, 1970 - Separation Time, Hours vs. Separated oil expressed as % of total sample by volume

Waste Stream Sampling Time	1	1.5	2	4.5	8.5	19
Start of Wasting	None	-	-	-	-	-
2 min.	22.5	25	25.4	26.2	27.5	32.5
5 min.	11.9	14.3	15.5	17.7	17.9	28.6
8 min.	16.3	18.6	18.6	20.9	20.9	30.2

Test No. 2 Feb. 19, 1970 - Separation Time, Hours vs. Separated oil expressed as % of total sample by volume

Waste Stream Sampling Time	1	1.5	2	2.5	3	18
10 sec.	4.8	6	7.1	83	9.5	-
1 min.	None	1.2	-	-	-	1.2
2 min.	"	1.2	-	-	-	1.2
3 min.	"	1.2	-	-	-	1.2
5 min.	"	1.0	-	-	-	1.0
7 min.	"	1.0	-	-	-	1.0
9 min.	"	1.0	-	-	-	1.0
13 min.	53.1	-	-	-	-	59.2

Test No. 3 Feb. 20, 1970 - Separation Time, Hours vs. Separated oil expressed as % of total sample by volume

Waste Stream Sampling Time	1	1.5	3
10 sec.	40	41.4	41.4
2 min.	40.5	-	43.2
3 min., 30 sec.	59.5	60	60.8

reservoir to agitate the contents and maintain the insoluble lime compounds in suspension. Low pressure steam, 15 psig or less, is satisfactory for use in the overall ammonia still operation. The consumption of steam usually ranges from 1.5 to 3 lb. per gal. of ammonia liquor distilled[4]. The ammonia-free liquor (termed "still waste") leaves the base of the fixed leg through a trap. The CF&I Plant has available two stripping towers (ammonia stills) which operate in parallel and are designed for removing ammonia from excess ammonia flushing liquor.

The lime leg has not been used for twenty years. Operation of the free leg was discontinued in 1963 due to high maintenance cost and low demand for ammonia. One of the stills was reconditioned in 1969 and operated without using the lime leg. Operation of the full size still, as determined through the pilot scale stripping tests, was strictly a matter of maintaining a top temperature of 93-95°C with 15 psi steam.

The vent gas from the still is returned to the coke oven gas upstream of the primary cooler.

Partially plugged lines limited the amount of flushing liquor that could be treated; however, flows of 25-30 gpm were measured by keeping a record on the storage tank level. Grab samples of new flushing liquor and still effluent were obtained and immediately analyzed for ammonia nitrogen, phenol, cyanide, chloride, COD, sulfate and total alkalinity.

Tables 65 and 66 show results of test runs conducted on the plant ammonia still. These results confirm the data obtained in the pilot study. When top temperatures were above 93°C, ammonia reductions of 80-90% were obtained. Reductions in the 50-60% range were obtained when top temperatures declined below 91°C.

The pH is an excellent indicator of operating efficiency of the still. In those tests where good ammonia reductions were obtained the pH of the effluent stream was approximately 1-1.5 units below that of the influent liquor. In the case of ammonia reduction, only 0-0.3 pH reduction was noted.

Although excellent cyanide removal was obtained at the higher top temperatures, removal of cyanide in this manner should be discounted because the stripped gas is returned to the gas stream and the stripped cyanide would appear in the final cooler water. Steam stripping of the free ammonia from the excess flushing liquor in the existing ammonia stills appears to be the best method of handling this pollutant. The stripped ammonia will be returned to the gas stream where it will be picked up at the saturators and converted to ammonium sulfate slurry. Controls will be installed to automatically control and monitor the top temperature of the still. The pH will be used to confirm ammonia reductions through the system. Excess flushing liquor to be treated will vary from 40 to 60 gpm dependent upon operating levels. The ammonia resulting from the stripping operation can be treated readily in the existing adsorber. Since sufficient low pressure exhaust steam is available for operation of the stills the principal operating cost will be consumption of 6-8% more acid.

TABLE 65

Results of Operation of Plant Ammonia Still

	Temp Deg. °C	pH	NH_3	Phenol C_6H_5OH	Cyanide CN	Sulfur Na_2SO_4	Chloride NaCl	COD	Total Alk. $CaCO_3$
11/14/69									
8:00 AM									
In	-	-	2074	-	-	-	-	-	-
Out	-	-	467	-	-	-	-	-	-
9:00 AM									
In	-	-	1870	-	-	-	-	-	-
Out	-	-	272	-	-	-	-	-	-
10:00 AM									
In	-	-	1870	-	-	-	-	-	-
Out	-	-	238	-	-	-	-	-	-
1:00 PM									
In	-	-	2006	-	-	-	-	-	-
Out	-	-	340	-	-	-	-	-	-
2:00 PM									
In	-	-	1904	-	-	-	-		-
Out	-	-	289	-	-	-	-	-	-
12/3/69									
8:00 AM									
In	64	8.8		-	-		-		
Out	85.5	8.5	1615	-		-			
9:00 AM									
In	65	8.6	2329	1280	47.7			7884	
Out	88.5	8.3	1564	1100	28.7		-	6894	
10:00 AM									
In	65	8.5	2312	-	-				
Out	88.5	8.4	1972						-
11:00 AM									
In	66	8.5	2057			-	-		
Out	88.5	8.3	1870	-	-	-	-		-
12:00 PM									
In	66	8.5	2261	1190	51	-	-	8091	-
Out	89	8.2	1649	1110	37	-	-	7389	-
1:00 PM									
In	66	8.6	2261	-	-	-	-	-	-
Out	87	8.3	1904	-	-	-	-	-	-
4:00 PM									
In	-	-	2108	-	-	-	-	-	-
Out	-	-	229	-	-	-	-	-	-
5:30 PM									
In	-	-	2074	-	-	-		-	-
Out	-	-	340	-	-	-			-
7:00 PM									
In	-	-	2006	-	-		-		-
Out	-	-	286	-	-		-		
8:30 PM									
In	-	-	2006	-	-	-	-	-	-
Out	-	-	292	-	-	-		-	-
Aver.									
In	65.2	8.6	2157	1235	49.4	497	304	7988	5300
Out	87.8	8.4	1172	1105	32.8	325	332	7142	3900

TABLE 66

Results of Operation of Plant Ammonia Still

	Temp Deg. °C	pH	NH_3	Phenol C_6H_5OH	Cyanide CN	Sulfur Na_2SO_4	Chloride NaCl	COD	Total Alk. $CaCO_3$
12/4/69									
8:00 AM									
In	65	8.7	2125	-	-	-	-	-	-
Out	91.5	7.2	187	-	-	-	-	-	-
9:00 AM									
In	65	8.6	2006	1270	45		-	7284	-
Out	91	7.5	340	900	3.9	-	-	6157	-
10:00 AM									
In	65.5	8.6	1972	-	-	-		-	-
Out	90	7.2	281	-	-	-	-		-
11:00 AM									
In	65.5	8.5	1931	-	-	-	-	-	-
Out	91	7.1	290	-	-	-	-	-	-
12:00 PM									
In	66	8.4	2040	1240	51.6	-	-	8294	-
Out	91	7.2	323	910	4.3	-	-	6085	-
1:00 PM									
In	65.5	8.5	2040	-	-	-	-	-	-
Out	91	7.3	306	-	-	-	-	-	-
3:00 PM									
In	-	-	2006	-	-	-	-	-	-
Out	-	-	272	-	-	-	-		-
4:30 PM									
In	-	-	1972	-	-	-	-	-	-
Out	-	-	286	-	-	-	-	-	-
6:00 PM									
In	-	-	1972	-	-	-	-	-	-
Out	-	-	272	-	-	-	-		-
7:30 PM									
In	-	-	2006	-	-	-	-	-	-
Out	-	-	272		-	-	-		-
Aver.									
In	65.4	8.6	2007	1255	48.3	552	170	8059	5000
Out	90.9	7.2	282	905	4.1	486	150	6121	-
12/5/69									
4:00 PM									
In	-	-	2023	-	-	-	-	-	
Out	-	-	279	-	-	-	-	-	-
5:30 PM									
In	-	-	1981	-	-	-	-	-	-
Out	-	-	286	-	-	-	-	-	-
7:00 PM									
In	-	-	1989	-	-	-	-	-	-
Out	-	-	272	-	-	-	-	-	-
8:30 PM									
In	-	-	2006	-	-	-	-	-	-
Out	-	-	-	-	-	-	-	-	-

STUDY OF OTHER TREATMENT METHODS

Other methods of treating coke plant wastes considered, but not tested were: 1) use of pure oxygen to improve oxygen input to activated sludge systems, 2) chemical oxidation, 3) electrodialysis, and 4) reverse osmosis.

Pure oxygen was considered for use in treating coke plant wastes by the activated sludge process. It is possible that a better-settling sludge would result from pure oxygen usage as opposed to air. This advantage would have to be weighed against the higher cost of oxygen purchase and storage.

Since removal of phenol is essentially an oxidation process, considerations have been given to inorganic oxidation techniques such as the application of chlorine, chlorine dioxide, and ozone. Direct chlorination of phenolic compounds may result in some destruction of phenol. However, formation of chlorinated phenolics introduces the possibility of taste and odor problems. Chlorine dioxide is a stronger oxidizing agent. The cost of this material and the difficulties in handling have prevented its application in most industries. Ozone is a much better candidate for oxidation of phenols because of the absence of side reactions and the high degree of phenol removal. Ozone consumption, however, is high and capital expenditures and high power costs are factors which have proved to be a major hurdle for usage in a number of instances.

Since the electrodialysis process is dependent on ionic charges of the chemicals to be removed, the process would have questionable application in a coke plant. While the ammonia and cyanide are both monovalent, the size of these ions are such that they can readily be transported across the membranes. The effective pore size of the ion exchange membranes is normally about 30 angstroms, although tighter materials, down to 10-15 angstroms, are available. Removal of phenol by means of an electrodialysis process would be virtually impossible because of the lack of electrical charges and also because of the possibility of high molecular weight phenolic compounds being present which would have a tendency to foul the ion exchange membranes.

Since the reverse osmosis is essentially a filtration process involving separation of dissolved molecules, pretreatment involving removal of suspended or colloidal matter is an essential requirement. Fouling of the membrane is again a major problem.

Most of the research has been conducted with cellulose acetate membranes which have been found to reduce the total dissolved solids level to acceptable values. However, it is known that celluose acetate membranes are permeable to ammonia. Phenols may also be passed by cellulose acetate membranes as indicated by actual laboratory studies[11]. It is doubtful that ammonia and phenol compounds would be removed by the membranes which are available commercially. Considerable research would be necessary in order to establish degree of removal and feasibility of a reverse osmosis process for coke plant waste treatment.

SOLID WASTE INVESTIGATION

The coal washery at the Pueblo Plant of CF&I Steel Corporation processes approximately 4,000 tons per 10 hour day of raw coal and operates seven days per week. The two major problem areas studied were the capacity limitations of the fine coal recovery system and the disposal of approximately 1,000 tons per day of waste solids washed from the coal.

An overall material balance was made on process solids and water flow through the existing coal washing system. Figure 1 (page 8) contains sampling point designations used in the material balance.

Grab samples were taken for this study. Test procedures used were those methods used by chemists of the U. S. Steel Corporation as outlined in "Sampling and Analyses of Coal, Coke and By-Products", 3rd Edition, copyright 1929. The samples were analyzed for percent total solids, percent moisture, percent ash by weight, percent volatile matter, BTU release, pyrometric cone equivalent for fusion point temperature, and volume reduction from burning. Screen analyses were also run to determine the particle size distribution of the solids.

Water flow measurements were obtained from pump design data, percent moisture in the grab samples, and a v-notch weir installed in the trench handling the waste effluent from the washery. Data are shown in Tables 103-106.

The material balance tables indicate that the washery is operating far below desired efficiency, especially with regard to coal recovery and residual slime. Table 107 is a compilation of important operating criteria for the mid 1960's and 1969.

The tables indicate that excessive coal is lost in the fine coal recovery section. Approximately 3 to 7 tons per hour (TPH) is leaving with the waste water. Ash content of the various coal fractions is also high, particularly in the fine coal fractions, which indicates residual dirt. Current high washery rates are also overloading the overall washery process with the result that the whole coal-refuse separation is no longer as efficient as originally designed.

The washery is designed to handle 300 TPH of raw coal. Factors such as more efficient labor scheduling and an increased demand for coke over the past decade have necessitated raising the washery throughput to its present level of approximately 400 TPH. There has also been a significant change in the raw coal consistency over this period (1959-1969) as shown in Tables 107 and 108.

TABLE 67 Process Analyses

Description	Total Solids % By Weight CF&I	Total Solids Normal Range CF&I	% Ash in Solids by Weight	% Volatiles by Wt. 9/15	% Volatiles by Wt. 10/7	% Volatiles by Wt. 10/23	BTU Release BTU/Gram 9/15	Volume Reduction From Burning	Melting Temp. of Ash ^{o}F
Washing Jig Large Refuse	87.0	85-95	81.0	13.2		14.9	582	12%	1521
Washing Jig Medium Refuse	81.8	80-90	71.8	17.9		17.3	2,657	14%	1491
Washing Jig Small Refuse	72.7	70-80	64.9	20.1		20.9	3.188	21.6%	1353
To R&S Flotation Unit	16.1	15-30	21.5			34.2			
Dewatering Screens Underflow	13.7	10-20	20.4			34.7			
To Decco Flotation Unit	26.7	15-30	25.8			33.5			
Fine Coal From R&S Flotation Unit	31.4	25-35	8.5	39.6		37.1			
Fine Coal From Decco Flotation Unit	34.0	25-35	8.3			37.3			
Overflow Waste	--	5-15	--	25.9	22.4		5,920	29%	1281
Tailing Screen Influent	36.6	30-50	38.9	30.2	27.2	30.5			

TABLE 68 Process Analyses

Description	Total Solids % By Weight CF&I	Normal Range CF&I	% Ash in Solids by Weight	% Volatiles by Wt. 9/15	10/7	10/23	BTU Release BTU/Gram 9/15	Volume Reduction From Burning	Melting Temp of Ash °F
Tailing Screen Refuse	56.9	50-60	21.3	40.6	21.9	35.7	13,485	76%	1215
Tailings from R&S Flotation Unit	12.0	5-15	30.0	26.2		33.3			
Tailings from Decco Flotation Unit	4.2	5-15	58.0	39.6		34.7			
Fine Coal Cake	72.4	65-75	--	37.6					
Waste Water Effluent	12.0	5-15	42.2	31.3		28.5			
Vacuum Filter Filtrate	8.4	2-5	13.8			36.1			
Recirculated Wash Water to Jigs	13.5	10-15	38.0			30.2			
Small Coal Dryer Feed	50.5	45-55	19.5			36.5			
Coal Froth	36.1	25-35	12.0			36.2			
Combined Flotation Tailings	14.8	5-15	53.0			26.0			
Medium Coal Dryer Feed	76.8	60-80	7.5			39.6			

TABLE 69 Process Flow and Material Balance

DESCRIPTION	WATER		SOLIDS IN TPH			GRAND TOTAL IN TPH
	GPM	TPH	COAL	REFUSE	TOTAL	
Total Washed Coal	132	33	274.5	25.5	300	333
Large Coal	25	6.5	116	4	120	126.5
Medium Coal	16	4	54	6	60	64
Small Coal	23	5.8	65	10	75	80.8
Fine Coal	68	17	39.5	5.5	45	62
Total Jig Refuse	34	8.5	20	59	79	87.5
Tailing Screen Refuse	6	1.5	0.5	1.5	2	3.5
Waste Water Effluent	600	150	7	6	13	163
Total Raw Coal	84	21	302	92	394	415
Vacuum Filter Filtrate	350	87.5	7	1	8	95.5
Coal Froth	418	104.5	46.5	6.5	53	157.5
Flotation Tailings	606	151.5	7.5	7.5	15	166.5
To R&S Flotation Unit	416	104	16.5	3.5	20	124
To Decco Flotation Unit	608	152	37.5	10.5	48	200

TABLE 70 Process Flow and Material Balance

DESCRIPTION	WATER GPM	WATER TPH	SOLIDS IN TPH COAL	SOLIDS IN TPH REFUSE	SOLIDS IN TPH TOTAL	GRAND TOTAL IN TPH
Medium Coal Dryer Feed	98	24.5	70.5	9.5	80	104.5
Medium Coal Dryer Effluent	82	20.5	16.5	3.5	20	40.5
Flotation Unit Dilution Water	334	83.5	--	--	--	83.5
Small Coal Dryer Feed	21	5.2	95.5	19.5	115	120
Small Coal Dryer Effluent	254	63.5	30.5	9.5	40	103.5
Flotation Machine Dilution Water	--	--	--	--	--	--
Mill Water Make-Up	334	83.5	--	--	--	83.5
Recirculated Wash Water	4152	1038	100.5	61.6	162	1200
Dewatering Screens Underflow	4433	1108	196	81	277	1385
Makeup Water	354	88.5	--	--	--	88.5
Washing Jig Coal Feed	4590	1147	402.5	153.5	556	1703
Dewatering Screens Feed	4556	1139	382.5	94.5	477	1616

TABLE 71 OPERATING CRITERIA

Item	1965	1969
Percentage Ash in Washed Coal	7-7.5	8.5
Percentage Ash In Refuse	75-80	70.8
Percentage Coal Recovery	95-96	91.0
Percentage Refuse in Raw Coal	20-30	23.9

TABLE 72 RAW ALLEN COAL SCREEN TEST

COAL	PERCENTAGE OF TOTAL	
SIZE	1959	1969
2" through 1/4"	57	41
1/4" through 0.131"	18	20
0.131" through 0.0328"	15	24
Smaller than 0.0328"	10	15

TABLE 73 TYPICAL OPERATOR DATA FOR VACUUM FILTER

Item	
Fine Coal Dried by Vacuum Filter	45 TPH
Moisture Content of Dried Coal	27 percent by Wt.
Solids conc. in Fine Coal Slurry to Vacuum Filter	34 percent by Wt.
Filtrate Flow Rate	350 GPM
Filtrate Solids Concentration	8.5 percent by Wt.
Vacuum Filter Cycle Time	1.5 Min./Rev.
Level of Vacuum of Filter Discs	21.5 - in. Hg°
Air Required to Discharge Filter Cake	5.5 CFM/Sq. Ft.
Percent - 200 M in Fine Coal	
Vacuum Filter Cake Thickness	0.75 - in.
Aerofloc 550 Additive	015 lb./ton/coal slurry
Type filter cloth in Service	60 x 40 mesh, 304 SS

This shift in consistency has increased the quantity of coal fines in the raw coal. The combination of increased throughput and revised consistency has overloaded the fine coal recovery system.

The present fine coal recovery system reached final development in 1959 when the coal washery throughput was 300 TPH. At this rate the flotation equipment handled about 22 TPH total solids, of which 15 TPH was recoverable coal fines. The estimated fine coal loss was 0.5 TPH. At present, the washery is processing as much as 400 TPH. At this high rate, the flotation equipment processes approximately 68 TPH total solids, of which 45 TPH is recoverable coal fines. However, the fine coal recovery section is incapable of recovering all the coal fines present in the feed slurry at this increased rate. As a result, 3 to 7 TPH of usable coal fines are discharged and wasted down the sewer with the high ash tailings from the flotation cells. The principal malfunction in the coal fines recovery system is the vacuum filter, which simply cannot process all the coal froth produced by the flotation equipment at the higher feed rates.

The present fine coal flotation section consists of two DECO[a] flotation machines and one R&S[b] flotation unit. Installation of flotation equipment occured as follows: one DECO unit in 1957, the R&S unit in 1958, and the second DECO unit in 1959. This resulted in a total flotation system volume of 9000 gallons. Since that time no major process or mechanical modifications have been made to the flotation system.

The feed to the R&S flotation unit is the effluent slurry from the three centrifugal dryers handling medium coal from the dewatering screens. The feed to the DECO flotation machines is the effluent from the two centrifugal dryers handling the underflow from the six cyclone separators. The coal froth produced in all flotation equipment is delivered to the sump of the vacuum filter. Tailings from all flotation equipment are collected and pumped to six cyclone separators and then to the tailing screens for a final dewatering.

A solution of methyl isobutyl carbinol (MIBC) in diesel fuel is added to the flotation feed streams. MIBC acts as a frothing agent, while the diesel fuel promotes particle agglomeration.

[a] Denver Equipment Company, 1400 Seventeenth St., Denver, Colorado 80217

[b] Roberts & Schaeffer, 120 So. Riverside Plaza, Chicago, Illinois 60606

Criteria which must be considered in operating a coal flotation system are: (1) the size and shape of the particles, (2) the specific gravity of the material, (3) the pulp density, (4) the rate and uniformity of the feed, (5) the frothing reagents, (6) the rate of froth removal, (7) the pH of the water and pulp, (8) the presence of dissolved salts, and (9) the flotation circuit itself.

The most efficient recovery by flotation is obtained with a particle top size of about 48 mesh, although some 28 mesh and 10 mesh materials have been recovered effectively. When particles are smaller than 325 mesh, flotation becomes more difficult. The separation of coal fines from the slime is possible due to the difference in surface characteristics between the slime particles and coal particles and is not, strictly, a gravity separation method. Air agitation and frothing agents promote the formation of the coal froth from particles of all sizes.

In treating coal larger than 200 mesh, pulp density should be about 20% TSS by weight. When processing extreme fines smaller than 200 mesh, pulp dilution should be in the range 10-15% TSS by weight. In general, the finer the coal the more dilute the feed should be. Dilute conditions in the slurry feed to the flotation equipment imparts to the coal solids a maximum mobility in the cells. Conditions favoring the upward movement of the coal particles result in greater recovery of coal fines.

The type of coal and the methods by which it was mined influence the percentage of extreme fines produced. In general, automated mining produces more fines than manual mining. As coal fines increase in the consist, high ash refuse increases. Fine clay or other materials have a tendency to form slimes and cause trouble in the coal cleaning circuit. This fact must be considered in the design and operation of a flotation system.

Figure 77 is a description of the flotation process developments at CF&I over the past decade. The increased feed rate has reduced the effective treatment time from 15 to 8 minutes at the expense of slurry dilution. The flotation process at CF&I is approaching the minimum acceptable treatment time and is, thereby, losing the benefits of slurry dilution. If this situation continues, a point will be reached at which efficient coal separation is no longer possible.

Stream analyses taken within the flotation process show that coal is lost in the tailings because the flotation cells cannot be operated under conditions conducive to maximum coal recovery. In practice, insufficient frothing agent is added because the vacuum filter cannot handle the extra froth produced without overflowing the sump. The

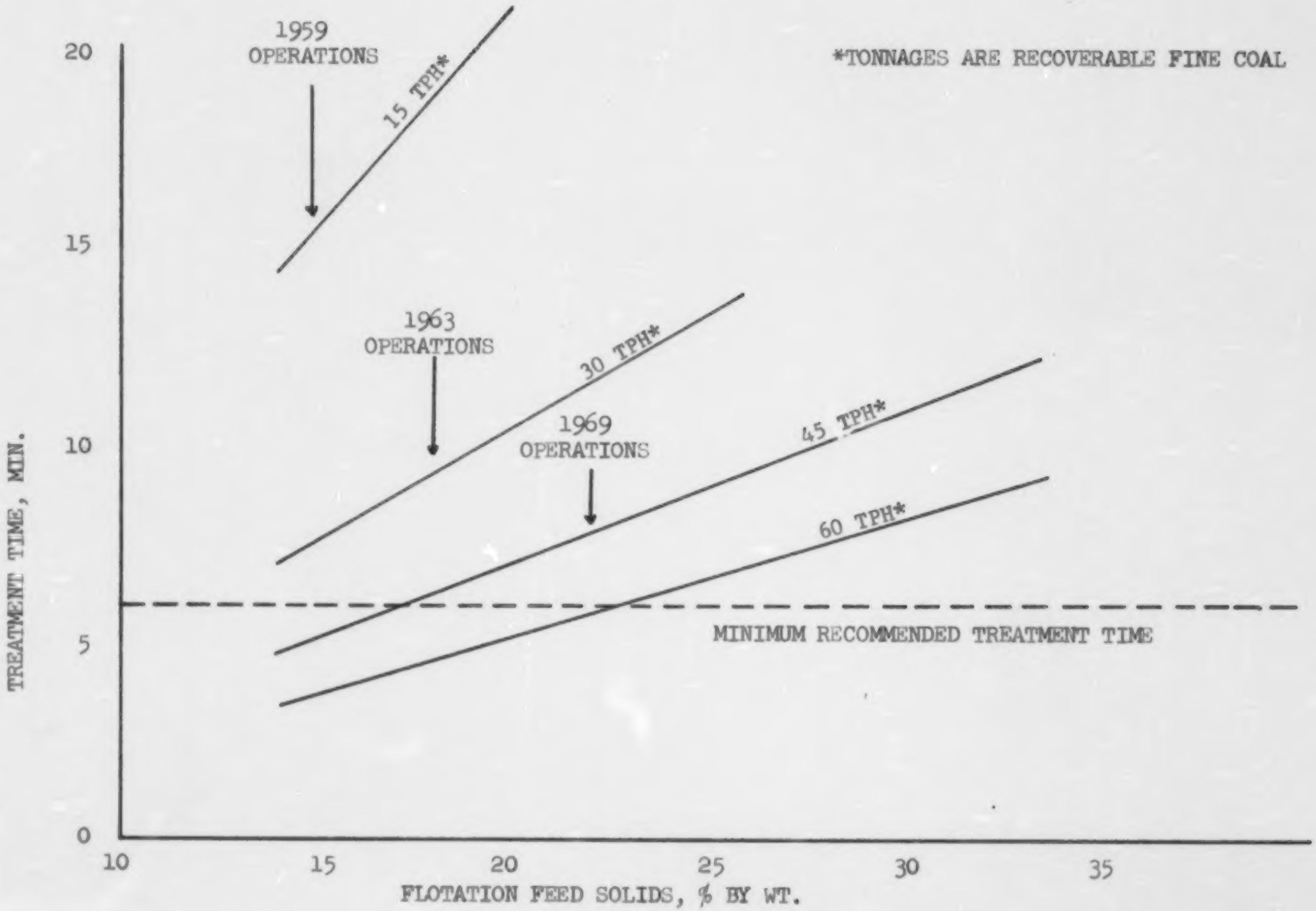


FIGURE 65. FLOTATION MACHINE TREATMENT TIME VS. SLURRY CONCENTRATION

correct dosage should be about one pound of agent per ton of dry coal solids. Calculations indicate only one-half of that amount is being added. This partly explains the large amount of fines (-200M) present in the tailings. In addition, the flotation equipment is operating under a suppressed condition because of insufficient air supply. Additional air would increase the amount of fine coal recovered, particularly the large particles. Inspection of tailings samples reveals considerable coal particles of 20M size trapped in the settled slime. If both these conditions could be corrected, at least 90 percent of the fine coal presently lost could be recovered.

The vacuum filter has five filter discs (750 square feet of filter area) and a 150 HP vacuum pump capable of pulling 22-inch Hg vacuum. The filter was installed in 1959 and since that time only minor modifications have been made to the filtering circuit. A backup filtrate pump and feed sump recirculation system were added in early 1960.

Filter feed is a mixture of froth from the flotation equipment and a flocculating agent, Aerofloc 550 (added to promote particle agglomeration). The filtrate extracted from the slurry feed is returned to the flotation equipment. Coal cake from the vacuum filter drops to a conveyor feeding the washed coal conveyor.

Any attempt to measure the quantitative effect of all variables influencing fine coal filtration would be virtually impossible. There are, however, several variables which can be measured to evaluate the filtration characteristics of fine coal. These include: (1) the solids concentration of the filter feed, (2) the filter cycle time, (3) the negative pressure differential across the cake, (4) the filter media, and (5) the particle sizing of the filter feed.

Each slurry has its own characteristic filtration curve (filtration rate vs. % solids). In general, a 28x0 mesh coal slurry feed exhibits a sharp increase in filtration rate above 35 percent solids. Controlling the solids concentration in the 45-55 percent range is optimum and requires less filtration area and lower operating expenditures.

As the coal particle size decreases, the feed solids concentration decreases. There is also a decrease in the maximum obtainable feed solids concentration. More filtration area is necessary and results are higher operating expenditures.

Cycle time of a continuous-vacuum coal filter is the time required for the filter to make one complete revolution, expressed as minutes per revolution (MPR). During the cycle, three phases of filter operation occur: (1) cake formation, (2) cake dewatering, and (3) cake discharge. Figure 78 is a plot of the formation rate of dry coal cake versus cycle time for a typical coal slurry. The slope of this curve is negative and is equal to -8.7 Lb/Hr/Ft^2/Min.

Filter cake moisture content is closely related to cycle time. Reduction of cycle time results in reduced cake moisture. Thicker cakes permit smooth and complete cake discharge and reduced maintenance costs.

To dewater the cake, it is necessary to provide sufficient vacuum to pull the required volume of atmosphere through the cake. The cake creates an air flow resistance which is recorded on a vacuum gauge as a negative pressure differential. It is this pressure differential which affects cake formation in the pickup zone and cake dewatering in the dry zone. For a filter cake of coal, a minimum air flow of 3 CFM/Ft^2 is required. Since most coal cakes are non-compressible, an air flow in excess of 3 CFM/Ft^2 has no deleterious effect. However, the benefits of the decrease in cake moisture content and the increase in cake formation rate become negligible beyond 6 CFM/Ft^2. To obtain minimum cake moistures and maximum cake formation rates for minus 28 mesh coal, a vacuum of at least 22-inch Hg is required together with optimum cycle time and solids concentration.

The use of the proper filter media is essential for effective operation (filtration rate, cake moisture content, and filtrate clarity). In addition, filter blinding (sealing of media openings resulting in thin cake because of low vacuum) and filter lifetime are important criteria. The most effective filter media today are stainless wire mesh cloth, saran, and polyethylene. Type 304 stainless steel 60x40 mesh wire cloth is the medium most widely employed on coal filters because of its long lifetime. The typical solids characteristics in the filtrate obtained with this cloth are 1.0-3.5% by weight concentration and minus 200 mesh.

To evaluate the vacuum filter performance, current operating data were acquired and is summarized in Table 109. The data definitely indicate inefficient filter operation while running at maximum capacity. The feed concentration of 34 percent solids is about normal; but a concentration less than that is possible if the flotation equipment is operated more efficiently. The filter cycle time is close to the absolute minimum as shown by the high cake formation rate, the abnormally high solids concentration in the filtrate (8.5%) and the high moisture content of the filtered cake. If a cycle time of 3 to 4 MPR were used, solids in the filtrate would be no more than 4% by weight and moisture content of the dried cake would be 20-25%. Drying air volume is

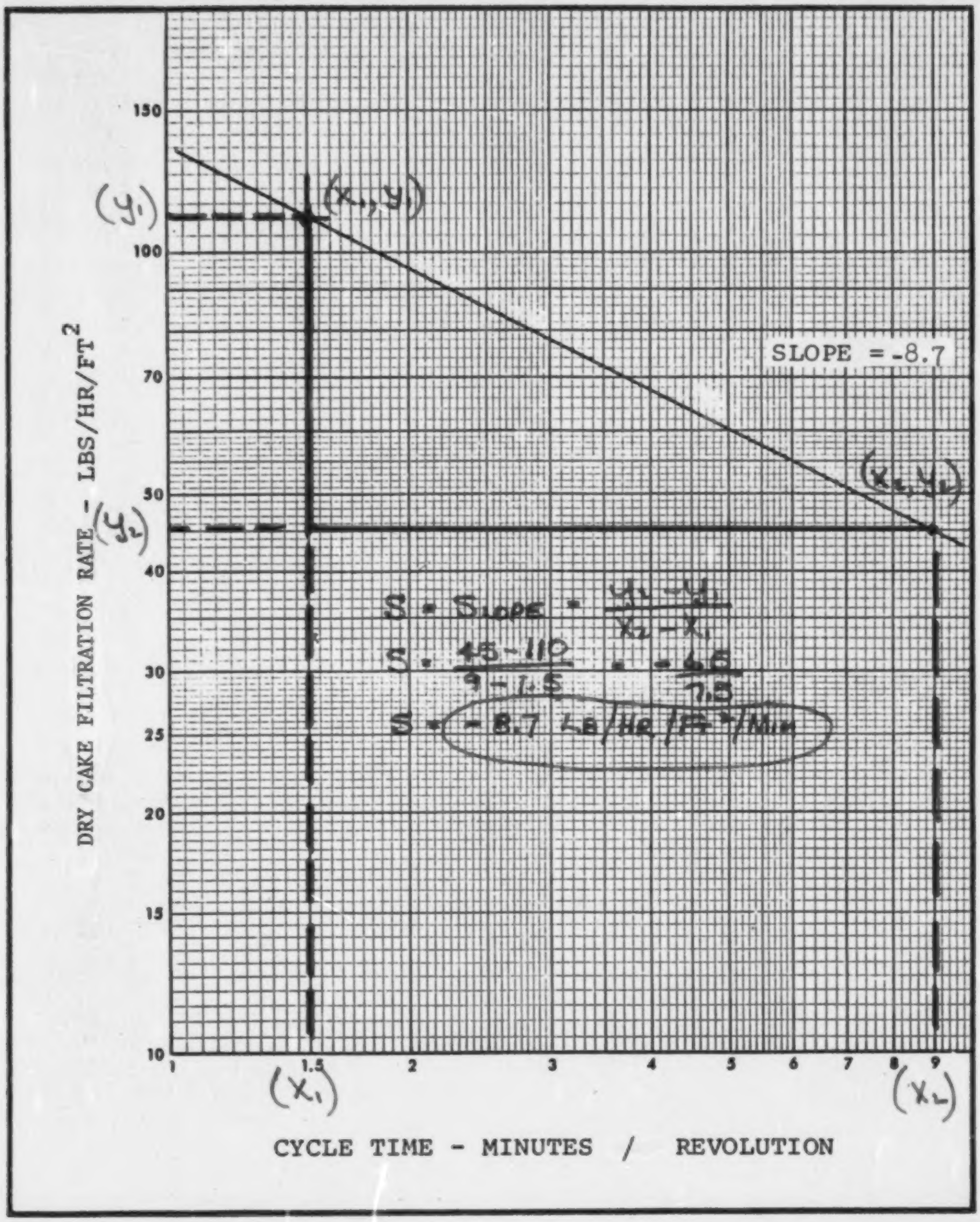


Figure 66. DRY CAKE FILTRATION RATE AS A FUNCTION OF CYCLE TIME FOR COAL AT CONSTANT OPERATING CONDITIONS AND SIZE CONSIST

more than adequate, but excessive filter speed negates the beneficial effect of the extra air. Cake discharge has presented no problem, but a thicker cake would reduce the permeability of the coal fines. The consistency of the slurry in the sump could also be improved (thickened) with the usage of additional Aerofloc 550. The current Aerofloc 550 dosage of 0.5 lb/ton of dry solids is about half that recommended for maintaining a satisfactory slurry for filtration.

The present filter was designed to handle 15 TPH of wet coal fines having a size range of 20-100 mesh. However, variable speed control of the discs in the range 7.6 to 1.2 MPR results in a filter capacity of 15 to 45 TPH. Optimum running conditions for the filter are 30 TPH and 3.0 MPR. Over the past decade, however, various factors such as increased washery throughput and changing coal consist have raised the filtration requirements to approximately 53 TPH. At present, the filter is running at maximum speed and is still unable to process all of the froth supplied by the flotation equipment. Occasionally an excess of frothing agent is accidentally fed to the flotation section, creating a froth load which the filter cannot handle. When this happens, the slurry tank overflows and coal escapes to the sewer system via either a recirculation buildup in the tank or the pump sump on the recycle overflow. This amounts to several tons per hour loss until the deficiencies are corrected. In addition to this, a more severe coal loss (3 to 7 TPH) occurs in the flotation section as a result of preventing filter overload by running these units below capacity. Reduced dosages of frothing agent and the inability to use outside air for additional agitation in the cells contribute to inefficient flotation operations and increased coal loss to the flotation tailings. In the current process, the coal leaving with the tailings is wasted to the solid refuse or to the waste water effluent stream.

The rotary disc filter handling the minus 20 mesh coal concentrates is the main problem in the coal fines recovery system. The filter is unable to process the maximum load of coal fines. When this problem is compounded by unscheduled filter breakdown, the entire washery must be shut down.

Fine Coal Recovery

A significant portion of this study was concerned with methods of improving the coal fines recovery in the Flotation and Coal Filter Systems. Results of the study indicated that increased efficiency is possible in the flotation system and that an additional filter in the filtration system would enhance the capability of that system.

This section presents a discussion of these modifications and a schematic diagram (Figure 79) of the proposed improvements. Estimated costs of the recommended modifications are also provided.

Over the past decade, significant strides have been made in flotation technology which have resulted in new and improved equipment. The Denver "Sub-A" flotation machine in use at CF&I has been outdated by a newer model called the "D-R" machine. The advantages of the "D-R" unit include: (1) better pulp recirculation with no stratification, (2) elimination of impeller sand choking, (3) improved aeration, and (4) better power efficiency.

The study revealed that the majority of the fine coal lost in the flotation tailings originated in the Denver cells. Only a small portion came from the R&S cell. The R&S cell performs a better coal-slime separation because of the type of feed handled and because aeration is used. In addition the Denver cells operate under a suppressed condition as previously explained, which further hinders flotation performance.

There are two approachs to improving the overall performance of flotation. The first is to replace the eight existing Denver "Sub-A" cells with eight new Denver "D-R" cells. With the installation of additional vacuum filters capable of accepting all of the flotation froth, approximately 90 to 95 percent of the coal fines presently lost in the flotation tailings can be recovered. The recommended new installation includes a 3000-gallon preconditioning tank for mixing the flotation feed with diesel fuel and MIBC, a pump to transfer the contents of the preconditioning tank to the flotation cells, and a 1900 CFM air compressor to supply flotation air.

The approximate cost these recommendations including equipment, installation, and engineering is given below:

ALL NEW DENVER CELLS

8 Denver "D-R" Cells w/Accessories, Installed	$50,000
Air Compressor, Installed	3,850
Preconditioning Tank and Pump, Installed	8,650
Piping	5,800
Electrical	3,450
Structural	6,800
Existing Cells Tearout	1,250
TOTAL	$79,800*

*Engineering costs are included in each of the above items.

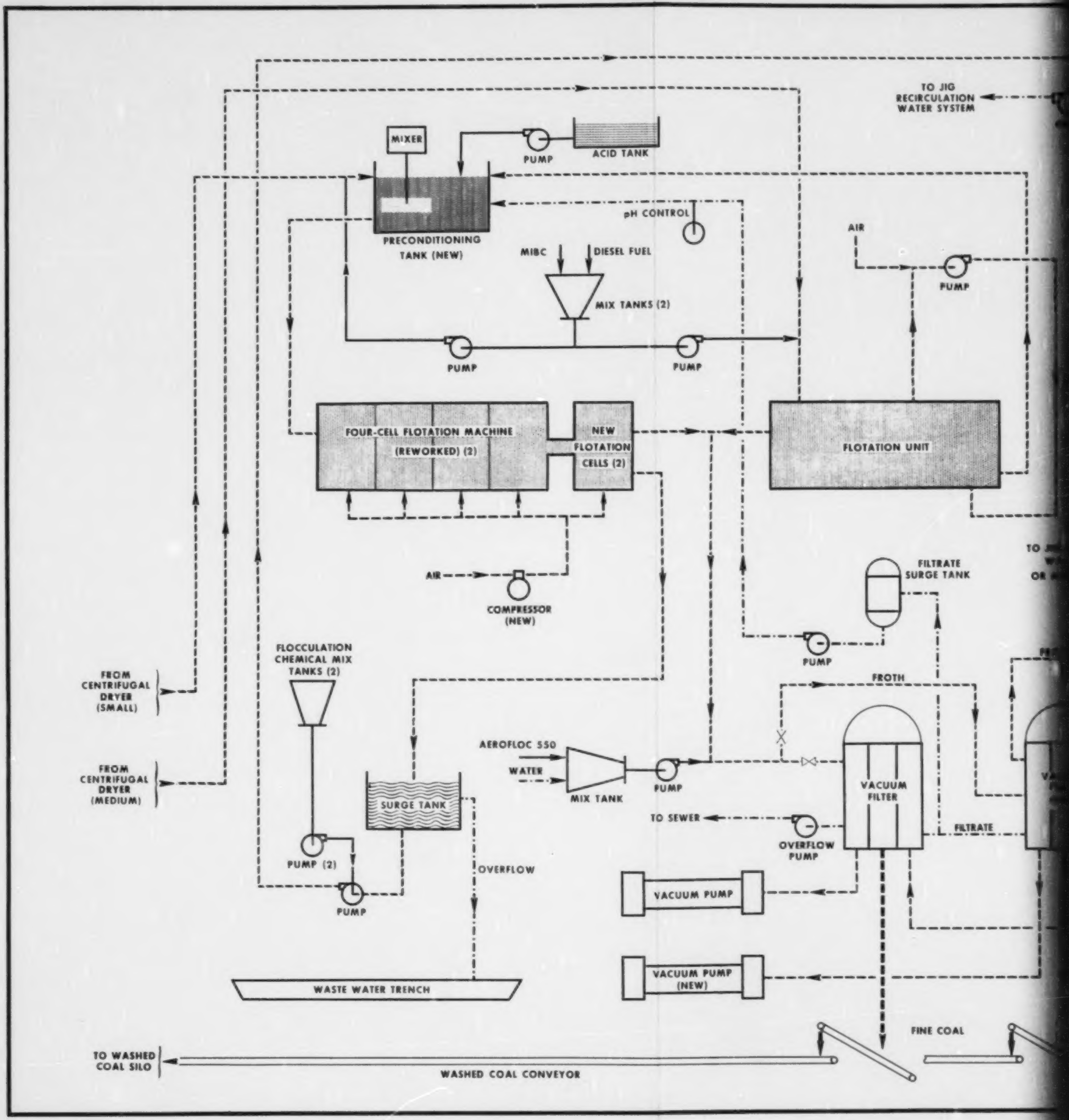
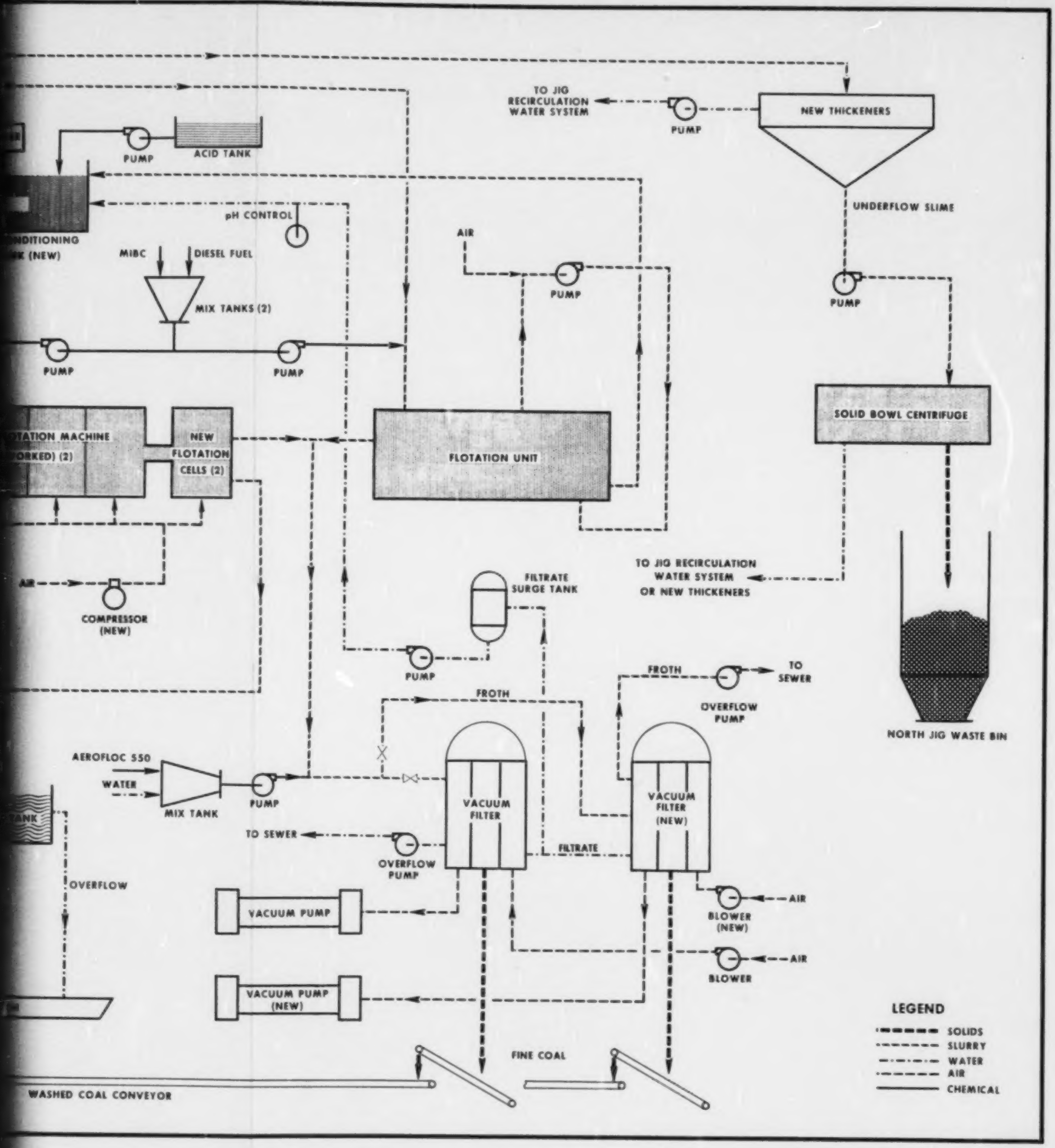

TO JIG RECIRCULATION WATER SYSTEM
MIXER
PUMP
ACID TANK
PRECONDITIONING TANK (NEW)
pH CONTROL
MIBC
DIESEL FUEL
MIX TANKS (2)
AIR
PUMP
PUMP
PUMP
FOUR-CELL FLOTATION MACHINE (REWORKED) (2)
NEW FLOTATION CELLS (2)
FLOTATION UNIT
AIR
COMPRESSOR (NEW)
FILTRATE SURGE TANK
PUMP
FLOCCULATION CHEMICAL MIX TANKS (2)
FROM CENTRIFUGAL DRYER (SMALL)
FROM CENTRIFUGAL DRYER (MEDIUM)
FROTH
AEROFLOC 550
WATER
MIX TANK
PUMP
SURGE TANK
VACUUM FILTER
TO SEWER
OVERFLOW PUMP
FILTRATE
PUMP (2)
PUMP
OVERFLOW
VACUUM PUMP
VACUUM PUMP (NEW)
WASTE WATER TRENCH
FINE COAL
TO WASHED COAL SILO
WASHED COAL CONVEYOR

Figure 67. FLOW DIAGRAM PROPOSED COAL WASHERY SYSTEM

A


TO JIG RECIRCULATION WATER SYSTEM
PUMP
NEW THICKENERS
PUMP
ACID TANK
ONDITIONING
NK (NEW)
pH CONTROL
UNDERFLOW SLIME
AIR
MIBC
DIESEL FUEL
MIX TANKS (2)
PUMP
PUMP
PUMP
PUMP
SOLID BOWL CENTRIFUGE
OTATION MACHINE
ORKED) (2)
NEW FLOTATION CELLS (2)
FLOTATION UNIT
TO JIG RECIRCULATION WATER SYSTEM OR NEW THICKENERS
FILTRATE SURGE TANK
AIR
COMPRESSOR (NEW)
PUMP
FROTH
FROTH
TO SEWER
OVERFLOW PUMP
NORTH JIG WASTE BIN
AEROFLOC 550
WATER
MIX TANK
PUMP
VACUUM FILTER
VACUUM FILTER (NEW)
TANK
TO SEWER
OVERFLOW PUMP
FILTRATE
OVERFLOW
VACUUM PUMP
BLOWER (NEW)
AIR
BLOWER
AIR
VACUUM PUMP (NEW)
LEGEND
SOLIDS
SLURRY
WATER
AIR
CHEMICAL
FINE COAL
WASHED COAL CONVEYOR

7. FLOW DIAGRAM PROPOSED COAL WASHERY SYSTEM

B

The advantages of this plan include: (1) use of the latest flotation equipment, (2) reduced cell maintenance, and (3) lower power and chemical consumption due to higher mechanical and separation efficiencies. The disadvantages of this approach are: (1) no extra capacity or spare equipment for operations flexibility, (2) no backup system to recover coal fines breaking through the Denver cells and R&S cell during process upsets, increased feed rates, or fine coal loading, (3) shutdown time for equipment changeover, and (4) capital cost.

The second approach to improving the performance of flotation is to convert the eight existing Denver "Sub-A" cells to the "D-R" configuration, add two additional "D-R" cells, and install a flotation feed preconditioning tank. This plan provides greater cost savings and operating flexibility than does the first plan.

The internal mechanisms of the existing Denver "Sub-A" flotation machines can be replaced with new equipment. When this is accomplished, the eight modified cells will achieve a level of performance and operating economy approaching that of the new "D-R" cells. Cell modifications include the following:

1. Replacement of impellers and drive mechanisms
2. Addition of new feed and discharge boxes
3. Vessel rework and addition of launders (troughs)

Although it is estimated that the eight modified Denver cells will remove approximately 90 percent of the coal fines in the feed slurry, a significant amount of fines will still be lost. In order to provide a complete and flexible arrangement, two new Denver "D-R" cells are required to process the combined tailings from both the R&S cell and the eight modified Denver Cells. The function of the two new cells will be to separate the residual flotable coal fines in the combined tailings, to protect the system from excessive coal losses in case of an operational upset, and to provide additional processing capacity.

As in the first plan, a 3000-gallon preconditioning tank is required to mix the flotation feed with the frothing agents. An additional pump is necessary to transfer the feed to the flotation cells.

The approximate cost of this plan, including equipment, installation, and engineering, is summarized on the following page.

MODIFIED DENVER CELLS

Modifications to 8 Existing Cells	$29,250
2 Denver Cells, Installed	9,200
Air Compressor, Installed	5,300
Preconditioning Tank and Pump, Installed	9,000
Piping	12,300
Electrical	3,250
Structural	6,100
Concrete	1,200
Tearout	1,400
TOTAL	$77,000*

*Engineering costs are included in each of the above items.

The major advantages of this approach are: (1) maximum fine coal recovery capability, (2) lower capital cost, and (3) minimal washery downtime for equipment installation and modification. Disadvantages inherent in this plan are: (1) higher operating costs, (2) increased maintenance, and (3) cell complexity.

All factors considered, however, the second plan employing the modified Denver "Sub-A" machines supplemented with two new units is recommended as the better method to improve flotation performance in the fine coal recovery system.

As previously discussed, the vacuum filter in the fine coal recovery circuit is the limiting factor in the system capacity. Operating data supports the fact that the existing vacuum filter is fully loaded, and sometimes overloaded. Under these conditions, filtration efficiency is low and the coal cake moisture is much higher than desirable. As previously stated, the filter cannot handle all the coal fines available from the flotation cells. For these reasons it is recommended that a second vacuum filter be installed (See Figure 79) which is similar in size and design, and is situated in parallel to the existing filter. The benefits derived from an additional filter are:

1. Provide all necessary fine coal filtering capability.
2. Provide extra capacity to handle surges, gradually increased washery feed rates, or changes in raw coal composition.
3. Provide backup insurance against total washery emergency shutdowns by having at least one filter in service at all times.

The cost of an additional filter, including vacuum pump, conveyors, tanks, installation, and engineering is estimated as follows:

NEW FINE COAL FILTER

Vacuum Filter, with all related equipment, installed	$88,800
Piping	5,600
Electrical	7,600
TOTAL	$102,000*

*Engineering costs are included in each of the above items.

This investment must be measured against the benefits derived (1) by recovering 3 to 7 TPH of fine coal presently lost, (2) greater filtering capacity to handle the anticipated increase in fine coal loads in the future, (3) a reduction in plant sewer maintenance, and (4) improved pollution control.

Disposal of Wastes

The coal washery generates two forms of waste: (1) a solid waste composed of shale, clay, and agglomerated pieces of coal and shale, and (2) a waste water slurry containing shale particles, clay, and coal fines.

The present tailings disposal system consists of six cyclone separators for crude solids separation followed by two tailings screens for refuse dewatering. The tailings screens are 48 inch Sweco[a] units, one of which has been in service for 16 years and the other for 10 years. Final arrangement of this disposal system was achieved sometime in the early 1960's.

The combined tailings from the flotation equipment are pumped to the cyclone separators where the tailings are concentrated. The tailings are then fed to the tailings screens where the solid refuse separates and is partially dewatered. The solid refuse from the screens passes to the north jig refuse bin where it is hauled by truck to the slag dump area. The effluent from the cyclone separators and the tailings screens flows to the waste water trench outside the washery building. The resulting stream (600 GPM) contains 4-12% total solids, of which 3-7 TPH is coal fines and the balance is high ash slime. The coal fines in the waste water represent about half of the total suspended solids in the waste stream. The trench feeds a collection sump connected to the coke plant sewer which exits eventually to Black Lake and thence to the Arkansas River.

To eliminate water pollution, the suspended solids must be removed from the waste water slurry. The present system of accomplishing this is inadequate.

[a] Southwestern Engineering Company, 6111 E. Bandini Boulevard, Los Angeles, California 90054

The rates presented below are based upon complete coverage (707 ft^2) of the basin with tubes.

Influent Waste TSS Concentration % By Wt.	Allowable Tube Overflow Rate (gpm/sq. ft.)	Influent Rate To Allowable/Basin (gpm)
2	1.0	500
4	0.62	330
6	0.40	210
8	0.25	132

The foregoing table reveals that the addition of tube settlers in the existing 30 ft. clarifier would be impractical because (1) the coal washery waste flow to be handled is 500-600 gpm, (2) the TSS concentration of the waste is 4-12% by weight, and (3) the usable surface area in the 30 ft. clarifier is 14% less than that for which these figures were based.

Sludge Dewatering

An experimental Eimco[a] belt-type vacuum filter was used to determine ability of this type filter to dewater the slimes in the clarifier underflow. The results presented in the Filtration and Sedimentation section (page 187) show that vacuum filtration of the slimes is not practical. This conclusion was confirmed by vacuum filter leaf tests conducted by Eimco in their Pallatine, Illinois laboratory.

Samples of concentrated washery waste placed in a small laboratory centrifuge indicated that it might be feasible to dewater the solids by this method. A 55 gallon drum sample was sent to the Bird Centrifuge Corporation[b] for a feasibility study. Initially the pH of the sample was adjusted to pH 6.5-7.0, and 8 mg/l Nalco D-2339 polyelectrolyte was added. Using a 6 inch solid-bowl continuous centrifuge, the solids recovery was 99% by weight and the cake moisture content was 40-45% by weight. While the moisture content is higher than desired, the centrifuge could be properly located to avoid problems in solids handling.

[a] The Eimco Corporation, 601 Broadway, Denver,
Colorado 80203

[b] Bird Machine Company, South Walpole,
Massachusetts 02071

Clarification

Filtration and sedimentation tests were conducted on the waste water from the coal washery. Results of these tests (see section "Filtration and Sedimentation Coal Washery", pages 154-188) indicate pH adjustment, polyelectrolyte addition, and a clarifier 80 ft. in diameter are required to produce an effluent which meets the requirements for re-use water in the jig wash water circuit.

Stearns-Roger conducted a study of the possibility of rehabilitating an existing 30 ft. diameter Dorr Thickener as an alternative to the afore mentioned scheme. Neptune Microfloc, Inc.[a] experimented with the installation of tube settlers in the thickener and chemical coagulation for the purpose of increasing detention time and producing acceptable clarification. Batch and continuous-flow testing were conducted on a sample of washery waste adjusted to pH 6.5 with pickle liquor. In the batch test, two settling columns were used (each 5½ inches in diameter and 28 inches high). One column was filled with raw waste. The other was filled with raw waste mixed with Nalco D-2339 polymer (8 mg/l). The settling curve indicated that the zone settling velocity of the waste was increased from 0.69 inches per minute to 3.5 inches per minute through the use of D-2339 polymer. The settled sludge volume was reduced by about 50% with the use of the polymer. The overflow rate for thickened waste was as high as 1 GPM per square foot of thickener surface. The corresponding underflow rate was as high as 25% of the thickener influent rate. The supernatant from quiescent settling of one hour contained 15 to 37 mg/l TSS.

In the continuous flow test tube settlers were used having a 6 foot water depth. The D-2339 polymer was fed at the inlet to the tube settler. The dosage was in the range 8-10 mg polymer/liter coal washery waste. The effluent suspended solids was in the range 61-64 mg/l TSS at an overflow rate of 1 GPM/square foot of tube settler surface. The underflow TSS was 10% by weight.

Neptune reported that the overflow rates listed below appeared to be feasible for the tube settlers tested. They suggested an adjustment of pH to 6.5-7.0 with spent pickle liquor followed by addition of Nalco D-2339 polyelectrolyte. However, the existing 30 ft. diameter clarifier contains only 500 square feet of surface area which is available for tube installation.

[a] Neptune Microfloc, Inc., 1965 Airport Road
Corvallis, Oregon

An 80 ft. diameter clarifier is the planned size. The dewatered slime (containing at least 55 percent solids) would be sent directly to the solid waste bin. The overflow from the clarifier would be returned to the jig wash water circuit for reuse. Control of the dissolved solids in the proposed system is maintained by fresh make-up water (184 gpm). The fresh water replaces approximately 184 gpm of water retained in the washed coal and refuse; washed coal 132 gpm, jig refuse 34 gpm, and 16 gpm in the dewatered slime from the centrifuge.

Centrifugation is the preferred method of dewatering clarifier underflow because the waste can be processed and monitored without any addition of manpower. In addition, this method results in no waste flow to the sewers. Adequate dissolved solids control is maintained through retention of water with the washed coal and refuse.

The estimated costs for the modifications and additions required to treat the waste slurry from the coal washery by clarification and centrifugation are listed below:

CONVERSION ITEM	COST
Install 80 ft. dia. Clarifier	$ 54,000
Install Centrifuge	68,400
Install Chemical Feeders W/Controls	18,900
Install Piping & Pump	18,900
Install Electrical	3,600
Tear Out Existing Equipment	5,000
Concrete and Structural Work	39,000
Total Estimated Conversion Costs	$208,500*

*Engineering costs are included in each of the above items. The above are preliminary costs; see Stearns-Roger report under Coal Preparation Plant Design Data Appendix J, pages 272-280.

The estimated costs to operate the centrifuge are:

ITEM	COST/DAY
Operators (2 Man Hrs./Day)	$12.00
Maintenance	7.00
Electric Power	12.50
Chemicals (Flocculation + Filter Aids)	38.00
Total Direct Operating Cost Per Day	$69.50

An alternate method of handling the underflow slime from the clarifier is pumping directly to drying beds. Instead of filtering the underflow from the clarifier which has already been treated with flocculating agents, this thickened stream of slime can be pumped directly to drying beds located approximately 2.5 miles south of the washery. This location is in a low marshy area between the slag dump and Salt Creek. Figure 80 depicts a site available for the beds. Three beds or basins would be required, each approximately 10 acres in size. The excavated dirt from each basin would be used to build basin earth dikes.

In the process of drying, it is possible for the top layer of slime to dry and seal off the portion underneath. If only one basin is used, this would result in a continued buildup (in alternating layers) of undried slime which would be difficult to handle during periodic bed clean-outs. Theoretical calculations indicate that 10 acres of surface area are sufficient to dry a 10-hour slime run from the washery, assuming an evaporation rate for the Pueblo area of 80 inches of water per year. However, to provide sufficient drying time as well as extra capacity for bed maintenance and cleaning, a minimum installation of 3 beds is recommended, each 10 acres in size and 5 feet deep.

The estimated costs for the modifications and additions required to handle the waste slurry by drying beds are:

CONVERSION ITEM	COST
Install 80 ft. dia. Clarifier	$ 54,000
Pump, 200 Gal/Min., Rubber Lined	2,800
Piping (12,000 Feet, 4 In., Schedule No. 40, Buried)	40,000
Electrical	13,000
Structural	9,000
Earthwork	64,000
Concrete	30,000
Insulation	2,000
Chemical Feeders W/Controls	15,000
Total Conversion Costs	$229,800*

*Engineering costs are included in each of the above items.

The estimated cost to operate the drying beds are:

ITEM	COST/DAY
Operators* (6 Man Hrs./Day)	$36.00
Maintenance (1/2 Man Hr./Day)	3.50
Electric Power	4.05
Chemicals (Flocculation Additions)	35.00
Total Direct Operating Costs Per Day	$78.55

*Majority of operator time is spent in resetting water flow between the beds and in bed cleanout.

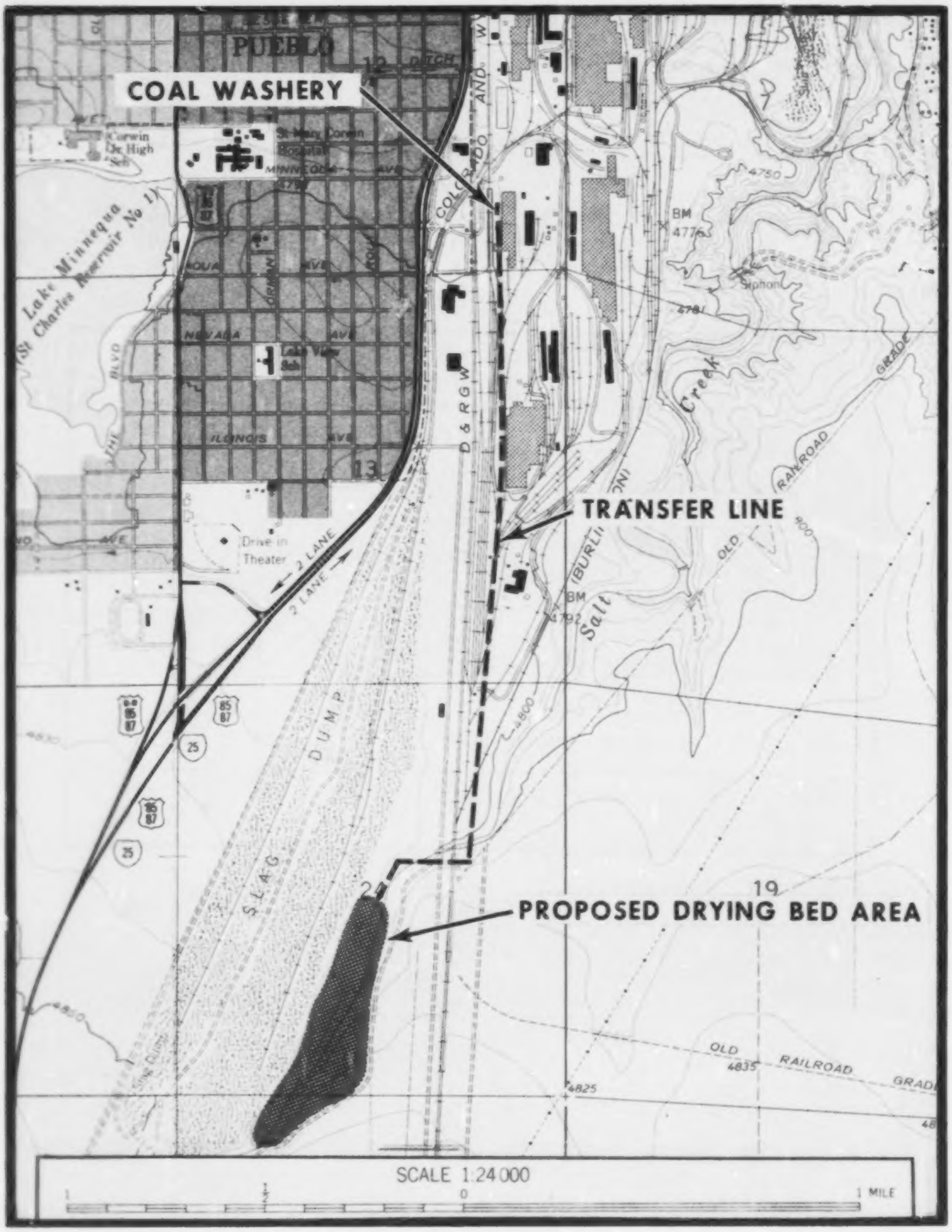

PUEBLO
COAL WASHERY
TRANSFER LINE
PROPOSED DRYING BED AREA
Lake Minnequa (St Charles Reservoir No 1)
St Mary Corwin Hospital
Corwin Jr High Sch
MINNEQUA AVE
NEVADA AVE
ILLINOIS AVE
Lake View Sch
Drive in Theater
2 LANE
COLORADO AND W
D & R G W
SLAG DUMP
Salt Creek
OLD RAILROAD GRADE
BM 4776
BM 4792
4750
4781
4800
4825
4835
4850
Siphon
12
13
19
25
85 87
SCALE 1:24 000
1
½
0
1 MILE

FIGURE 68. PROPOSED DRYING BED LOCATION

Marketing of Waste Refuse

The noncombustible refuse and slack from the coal washery (approximately 1,000 tons per day) are collected in the refuse bins and are hauled by truck to a landfill area. This refuse consists mainly of shale and inorganic matter having a size and gradation similar to that of the coal (plus 1½ inch to minus 100 mesh). Since the shale is sedimentary in nature, it exhibits poor structural qualities and contains a small strata of coal. The color of the refuse varies from gray to black.

The results of analyses performed on the refuse taken directly from the three elevators at the jigs are:

SIZE OF REFUSE	VOLATILES (% by Wt. at 950° C)	BTU RELEASE (one gram)	VOLUME REDUCTION BY BURNING (% by Wt.)	ASH MELTING TEMP (°C)	WEIGHT OF ASH (% of Refuse)
Large Refuse	13.2	582	12	1521	80.3
Medium Refuse	17.9	2657	14	1491	70.4
Small Refuse	20.1	3188	21.6	1353	68.1

The analyses indicate that the refuse contains coal, but the heat release is insufficient to support combustion (2142). The remainder of the refuse is silt and clay originating from sedimentary rock. The marketability of any such byproduct is dependent upon the chemical and physical properties of the material. It is unlikely that the chemical properties of the shale represent value in comparison to any other natural earth or solids. Due to the quantity available for disposal (1,000 tons per day) the refuse must have physical properties equal to sand, gravel, or slag in order to represent a marketable commodity. The refuse does not meet the requirements of coarse or fine aggregates that are marketable in the area. These requirements include color, limits on organic impurities, abrasive strength, and friable particles.

Another consideration in the marketability of the material is the hauling cost. The estimated commercial hauling costs for bulk dirt and rock materials are:

LENGTH OF HAUL	COST PER TON
2 Miles	$0.40
5 Miles	0.75
10 Miles	1.35
15 Miles	1.95
20 Miles	2.55

Since the physical properties of the washery refuse are poor in comparison to other fill materials available in the area, it is highly doubtful that a market can be established for the refuse. The present refuse dump is adequate for many years of refuse storage and is a well managed dump. Should a market for the refuse develop at any time, it can be sold directly from the dump area. However, no immediate market is foreseen.

Disposition of Waste Solids

The waste solids accumulated by either the filter method or by drying must be removed periodically. Upon removal, the waste solids must be disposed of in a manner that will meet the State antipollution standards. Two methods of disposal were evaluated: (1) incineration; and (2) landfill.

Incineration of Waste Solids

Incineration was considered as a possible solution for reducing the 1000 TPD of solid waste from the washery. Laboratory analyses indicated that a total of only 2142 BTU's per pound were released upon burning with a corresponding volume reduction of 16 percent.

This quantity of heat is insufficient for self-supporting combustion. Consequently, large amounts of externally-supplied heat would be required to vaporize the estimated 480,000 lbs/day of water associated with the waste and to elevate the remaining dry waste to combustion temperature.

Incineration is uneconomical and impractical because of the large external heat requirement, the low volume reduction obtained during incineration, and the high heat loss in the ash. In addition, an incinerator would require an elaborate and expensive fly ash collection and scrubbing system to minimize the possibility of air pollution.

Transfer to Landfill

The transfer of waste solids to a landfill area has been studied in detail. This method of waste disposal is more advantageous because sufficient land area is available. Four methods of waste transfer were evaluated: (1) pneumatic; (2) hydraulic; (3) rail; and (4) truck.

Pneumatic Transfer System

The pneumatic transfer system would require the grinding of approximately 1000 TPD of solid refuse to the proper size and the pneumatic transfer of the refuse to a dump two miles away. This would necessitate drying and pulverizing the waste solids prior to entering the transfer system. An estimated 480,000 lbs/day of water must be evaporated from the waste solids. Then the solids must be ground to less than 1/4 inch particle size. Approximately 1200 horsepower are required to operate the transfer system using a 12 inch diameter pipeline. Pneumatic lines extending long distances (over 500 feet) have relatively low efficiencies. This brings about a high horsepower requirement. A bag-filter dust collection system would be required at the dump site to trap the extremely fine particles and minimize air pollution. The filter system must be portable so as to distribute the waste evenly throughout the dump area. Additionally, the bag filter system would require a portable power supply and significant operator attention and maintenance.

The power, heat, and personnel requirements make this system economically unattractive.

Hydraulic Transfer of Landfill

The hydraulic transfer system would involve equipment necessary to handle approximately 1000 tons per day of solid waste and 400,000 gallons per day of liquid waste. The system would entail installation of conveying and grinding equipment and a new one-hundred ton solid waste bin at the washery. An 8 inch transfer line (approximately two miles in length) would have to be installed from the washery to the slag dump. The line would be laid on a grade and buried under road and railway at intersecting points. The solid waste would be ground to a particle size less than 1/2 inch, slurried with washery waste water, and pumped to the slag dump through the 8 inch transfer line. Makeup water amounting to 350 gallons per minute would be required for the system in order to maintain adequate transport velocities.

Modifications required at the slag dump for the storage of solids waste make this system extremely unattractive. The central area of the existing slag dump is surrounded on all sides by slag from the blast furnace operation. This material is currently sold, and portions are removed as required by the purchasers. Construction of a dammed area at this location is not considered practicable.

Alternate sites do not appear feasible. Possible sites along Salt Creek would have a limited life and would present problems of drainage, ground seepage, etc.

Rail Hauling to Landfill

A rail hauling system to dispose of 1000 tons per day of solid waste would require the installation of a new railroad spur at the coal washery and the relocation of existing tracks at the slag dump. Approximately 15 cars per day would be required for this routine transfer. Relocation of the railroad tracks at the slag dump would be required periodically to minimize the distance from rail unloading spot to the dump area.

The existing spur facilities near the coal washery (for raw coal unloading) are in heavy use. The insertion of additional cars for waste transfer would undoubtedly disrupt this operation. The hydrated solid waste would freeze during the winter months and would present problems in unloading at the dump. Since management of the dump area would become fairly complex, this method is not given further consideration.

Truck Hauling to Landfill

At the present time, solid wastes from the coal washery are being hauled by truck to a landfill area. Approximately 1,000 tons per day of solids are loaded and hauled to the existing slag dump located two miles southwest of the washery. A subcontractor performs this service at an approximate cost of $0.135 per ton. A 35-ton capacity truck handles approximately 90 tons per hour and makes a round trip every 20 to 25 minutes.

This method of transfer to landfill appears to present little or no pollution problems. The solid wastes are quite wet and present no dusting problems. On occasion, excess water in the wastes draining from the truck presents minor problems. This excess water could probably be minimized by the addition of flashing to the bucket conveyors at the coal jig, thereby deflecting water back into the jig.

This method of transferring the solid wastes to landfill has the advantages of minimal capital expenditures and low operating costs.

ENGINEERING DESIGN OF COAL WASHERY IMPROVEMENTS

A part of the overall project consisted of preparing an engineering design of a full scale facility which would eliminate the discharge of polluted water and would provide separation and recovery of the usable fine coal presently lost to the waste stream. Stearns-Roger Corporation, Denver, Colorado, was engaged to perform this work.

The scope of this engineering design consisted of preparing equipment specifications, obtaining and evaluating major equipment bids including selection of equipment for the layout design, and preparing mechanical layout drawings depicting the installation of equipment within the existing coal washery facility. The Stearns-Roger design conforms to the methods, equipment and approach of the Solid Waste Study section of this report. The major components of the engineering design are improvements in the flotation system including renovation of the eight existing flotation cells, addition of two new flotation cells, addition of a conditioning tank, and installation of a second fine coal filter.

The final effluent treatment system consists of a new 80 ft. diameter clarifier, a solid bowl centrifuge for sludge dewatering, and provision for recycling the water. Tube settlers for the clarifier are not planned for installation at this time. Should the future process require additional capacity, tube settlers will be reconsidered.

The estimated cost for the total installation as prepared by Stearns-Roger is as follows:

Electrical Installation	$ 10,807
Mechanical Equipment Installed	382,114
Civil-Structural Work Installed	51,170
TOTAL	$444,091

The total cost estimate determined by the engineering design is approximately 12% higher than the preliminary estimates developed in the Solid Waste Investigation and represents a more reliable figure.

Copies of Stearns-Rogers cost estimates and design drawings are included in the Appendix J, pages 272-285.

ACKNOWLEDGEMENTS

The support of the project by the Federal Water Quality Administration and the assistance provided by Mr. Fred Pfeffer, the Grant Project Officer, and Mr. William Lacey and Mr. Edward Dulaney of the Washington Office is acknowledged.

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Mr. William Boyer and Mr. Mark Bonomo headed the group from Stearns-Roger Corporation which provided the study, recommendations and design of the Coal Washery facilities.

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APPENDICES

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APPENDIX A

PILOT PLANT STRIPPING COLUMN (ASSEMBLY)

Feed System

The column influent was taken from a feed tank equipped with a drain and a feed pump. The influent to the column was metered by means of a ball type flow meter, capacity 0-6 gpm.

The feed pump required a drain opening, minimum size 3/4 inch, for removal of accumulated sludge and had a capacity of 6 gpm at a total discharge head of approximately 20 psig.

Stripping Column

The column was assembled from Items 1-4, as shown on Drawings PCD-142-CP3 and -CP4. A support bracket for the column was included to allow a clearance between the bottom drain valve and the ground level of approximately 42 inches.

The column was filled with $1\frac{1}{4}$ $ft.^3$ of 1 inch x 1 inch Raschig[a] rings.

As shown on Drawing PCD-142-CP3, Item 1 was furnished with a 3/4 inch x 3/4 inch screen for retention of the Raschig rings. The screen was welded to the inside of the column. In assembling the column, the screen was adjacent to Item 4.

The vent was extended to the outside of the building or to an area where fumes would not cause interference.

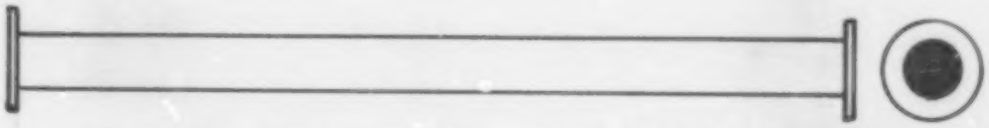
The still was operated with 100 lbs. of steam. The pressure was reduced manually in a throttle (3/4 inch globe valve) ahead of a 3/16 inch orifice. The orifice was installed between two standard flanges in the 3/4 inch steam line. Brass or stainless steel was used for an orifice plate. A pressure gauge was included in front of the orifice plate.

Uninterrupted pipe running approximately 12 inches was included on both sides of the orifice plate.

The drain connection was made from 1 inch standard schedule 40 pipe for the vertical legs. A discharge manifold consisting of five 1" outlet valves, $7\frac{1}{2}$" apart, extended from a level even with the bottom knuckle of the stripping column to a level approximately 30" higher.

[a] U. S. Stoneware, Inc., Akron, Ohio 44309

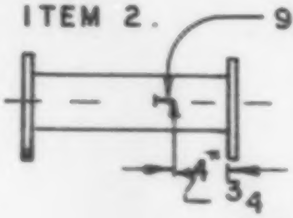
ITEM 1.



SCH. 40 STD. 6" PIPE 6FT. LONG
FLANGED ENDS.

3/4" 3/4"
SCREEN

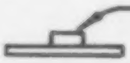
ITEM 2.



SCH. 40 STD. 6 PIPE 18" LONG
FLANGED ENDS WITH 3/4" PIPE
& 90° ELBOW ASSEMBLY.

3/4 PIPE & ELBOW ASSEMBLY
ELBOW ℄ ON PIPE ℄
3/4" PIPE 4 FROM END OF PIPE
EXTEND 2" FROM 6" PIPE WALL.

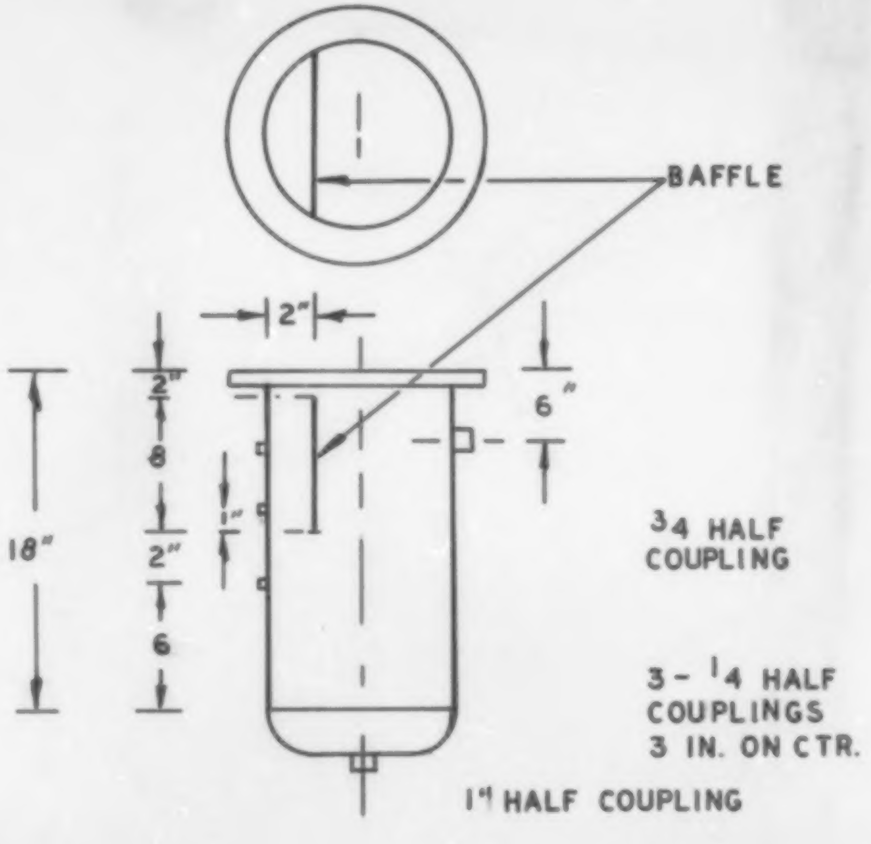
ITEM 3.


3" HALF COUPLING

STD. 6 FLANGE WITH 3 HALF
COUPLING IN CENTER

ITEMS 1, 2, & 3.
STRIPPING TOWER

DWG PCD 142 CP 3


BAFFLE
2"
2"
8
1"
18"
2"
6
6"
34 HALF COUPLING
3- 14 HALF COUPLINGS 3 IN. ON CTR.
1" HALF COUPLING

6" FLANGED PIPE CAPPED END

ITEM 4
STRIPPING COLUMN
DWG PCD-142-CP-4

APPENDIX B

PILOT PLANT STRIPPING COLUMN (OPERATION)

General:

The stripping column was designed for counter-current contact between ammonia liquor and a stripping gas such as steam or air. The liquor was introduced at the top of the column while the stripping gas was forced upward through the bed of Raschig[a] rings which provide the necessary contact between the gaseous and the liquid phases.

Calibration

1. Steam Nozzle

The steam flow rate was controlled by the manual valve in front of the steam nozzle. The nozzle was calibrated by condensing and collecting the steam discharged at approximately the same back pressure as was normally experienced during column operation. The calibration curve is shown in Figure 1. This figure relates steam discharge in pounds per minute to steam pressure in front of the nozzle in pounds per square inch gauge.

2. Water Flow Meter

The water meter had a scale indicating discharge in gpm of liquids having specific gravity values of approximately 1.0.

Calibration of the meter was necessary because of abnormal temperatures and different viscosities of the ammonia liquor. Calibration data for the flow meter is shown in Figure 2.

Operation

The feed tank was filled with the liquor to be stripped. Minimum time was lost between filling the tank and starting the stripping operation. The stripping operations began by adjusting the water flow meter to the desired level while the feed pump was running. Valve No. 2 was opened on the column discharge manifold. When water began discharging from the No. 2 valve, steam was admitted to the column at the desired rate, manifold valve No. 2 was closed, and the valve indicated on the operating table, Table 1, was opened.

[a] U. S. Stoneware, Inc., Akron, Ohio 44309

The column was allowed to run for a few minutes to permit accurate adjustments of steam and water flow rates. It was important that the switch from the No. 2 manifold valve to the valve indicated in Table 1 was performed rapidly so as to avoid blowing the water seal as a result of the increased steam pressure at the bottom of the column. Table 1 is based on approximate values of observed pressures at the increased rates of steam and water flow.
If the seal was blown, the column was restarted by temporarily shutting down the steam flow, opening manifold valve No. 2 until the bottom column was filled as indicated by discharge of water from valve No. 2, and then switching to the next higher valve on the manifold after introduction of steam.

It was important, also that the manifold discharge valve with the lowest possible number be used in order to avoid flooding the bottom of the column. The approximate water level at the bottom of the column was measured by opening the one-quarter inch side valve. It was desirable to maintain a water level below the middle half coupling.

The system was allowed to operate for approximately five minutes before any data was tabulated.

Lime Treatment

Fixed ammonia in the ammonia liquor was liberated by the addition of lime. It had been established that a pH of at least 11.2 would be required in order to reduce the fixed ammonia to levels below 50 mg/l.

Lime was added to the feed solution in the feed tank while the solution was agitated manually. Addition of lime was made by making a slurry of the lime in a 5 gallon can prior to blending with the feed solution itself. Direct addition of dry lime to the feed solution was avoided.

After reaching the desired pH level, the solution was allowed to stand for approximately 45 minutes - 1 hour in order to allow the lime sludge to settle. The pump intake was kept free of lime sludge.

Stripping of ammonia from lime-treated liquor was carried out as quickly as possible in order to avoid direct transfer of free ammonia from the feed tank to the atmosphere.

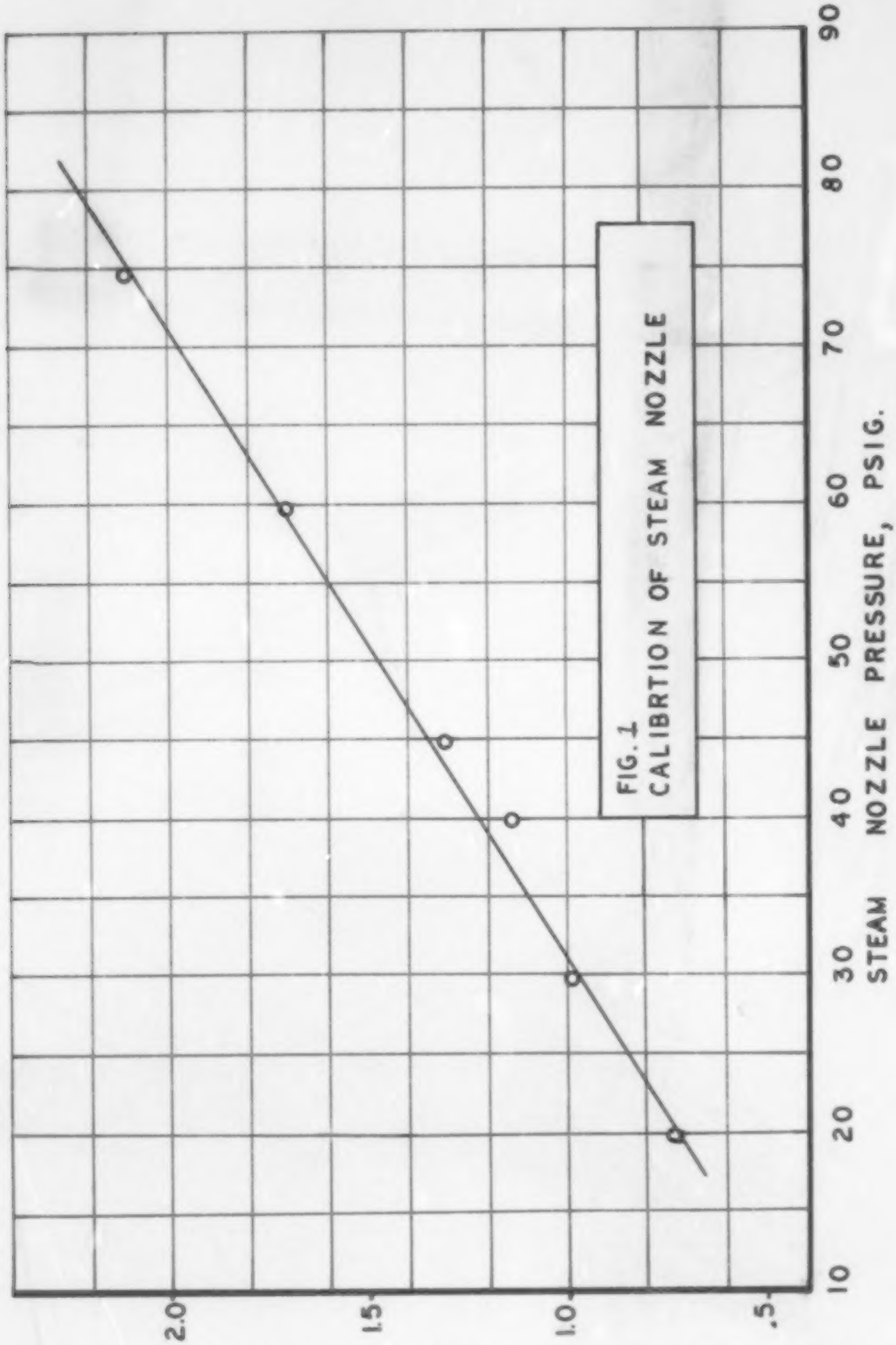


FIG. 1
CALIBRTION OF STEAM NOZZLE

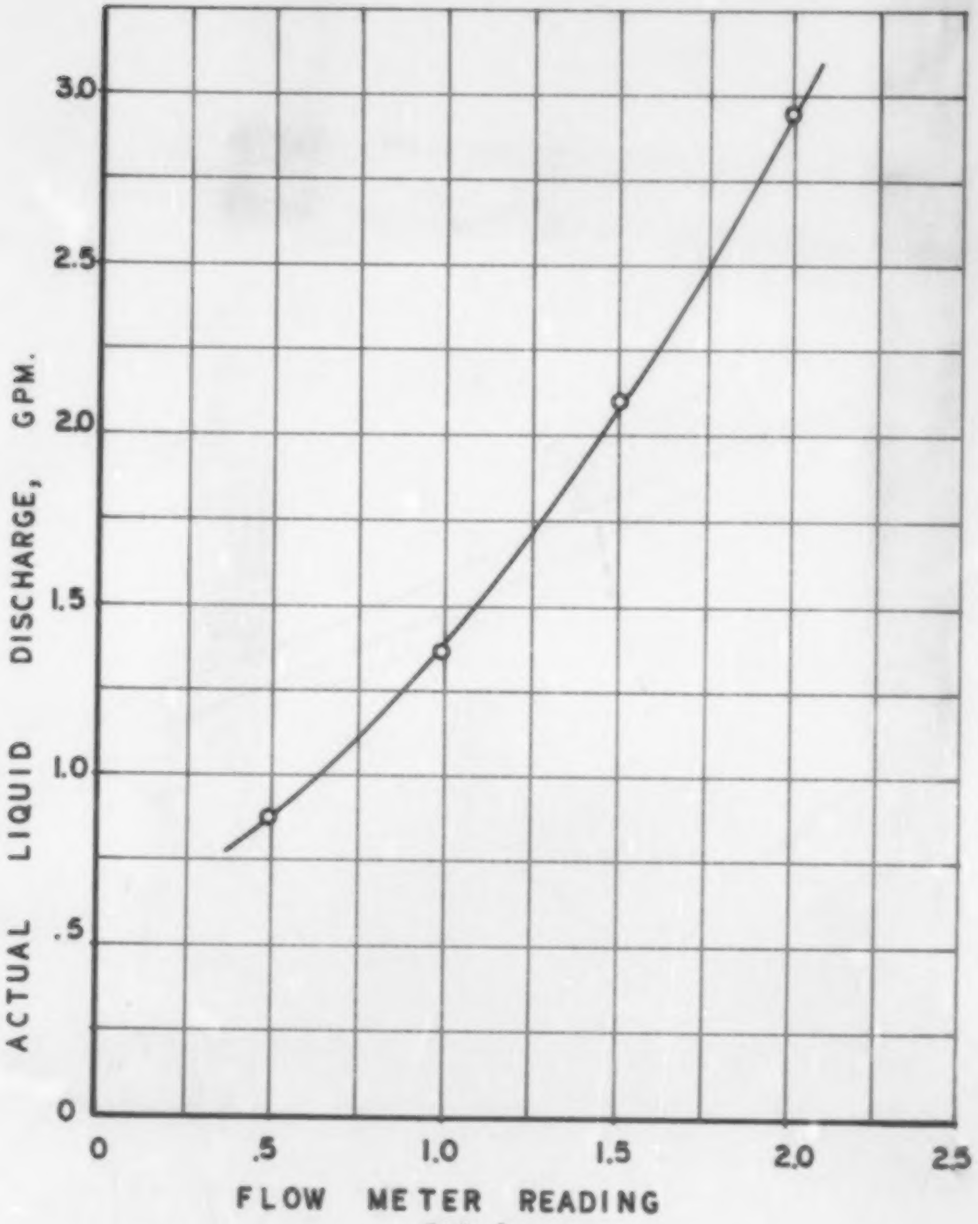


FIG. 2
CALIBRATION OF WATER FLOW METER

TABLE 1

Stripping Column Flow Rates and Approximate Water Seal Heights

Water Meter Reading GPM	Steam Nozzle Pressure, psig	Water Seal Pressure in Ft.	Use Manifold Discharge Valve Number
.5	30	.1	2
.5	45	.2	2
.5	60	.5	2
.5	80	.8	3
.5	100	1.25	3
1.0	30	.1	2
1.0	45	.2	2
1.0	60	.3	2
1.0	80	.6	3
1.0	100	.8	3
1.5	45	.3	2
1.5	60	.5	3
1.5	80	.8	3
1.5	95	1.5	4

APPENDIX C

PILOT PLANT TRICKLING FILTER (ASSEMBLY)

Feed System

The raw influent waste was added to a recycle sump through a float operated level control valve. A standard float valve may be used for this purpose; brass construction would be suitable. The valve was anchored to the recycle sump in order to permit proper adjustment of the liquid level.

In order to provide proper operation of the float valve, it was necessary to elevate the feed reservoir. Too little pressure on the float valve caused erratic operation and fluctuation of the liquid level in the recycle sump.

The elevated feed tank was operated from a primary feed tank located at ground level. The liquid can be recycled through the elevated feed tank, provided an overflow return line to the primary feed tank is included.

Tower Recycle System

The recycle sump was made from a 55 gallon drum, cut off approximately 12" from the bottom. The recycle pump was equipped with a flow meter capable of handling 0-2 gpm.

The recycle pump discharged into an overhead recycle reservoir equipped with a bottom outlet (1/2 inch NPT nipple with a 1/2 inch solenoid valve). The solenoid valve was actuated by a variable cycle automatic reset timer. Cycle time was adjustable between 1 and 5 minutes per cycle.

The bottom drain line from the recycle reservoir discharged to a distributor pan located at the top of the filter stack. The pan had a perforated bottom to allow proper distribution of the discharge from the recycle reservoir.

The recycle reservoir was assembled from a 5 gallon pan with a black iron nipple welded on at the bottom. The distributor pan was made from a plastic pan. Openings in the bottom were made by drilling 3/16 inch holes at the locations immediately above the channels of the SURFPACK modules in the tower. The holes were located in such a fashion that direct discharge from the distributor pan to the open channel through the SURFPACK was not possible. In other words, the holes were located immediately above the edge of the openings.

The 12 stack SURFPACK tower were made from sheet metal. The inside dimensions were 20 inches x 20 inches x 23 feet. Proper bracing was included to make the tower rigid for guy-wire support.

The tower was supported from the bottom on a wooden scaffold, approximately 3 feet above the ground. The scaffold was designed for positioning and support of one pan and a trough to be operated as follows.

The recycle collection trough was located in a slightly slanted position and discharged to the recycle sump. The trough was positioned as close to the ground as possible and provided gravity discharge to the recycle sump.

The effluent collection trough was located approximately 6 inches below the bottom end of the tower and was sufficiently large to catch all of the water from the tower. A 24 inch square pan, approximately 6 inches deep, with a bottom discharge spout (1/2 inch pipe) was suitable for this purpose. The effluent collection trough was supported on horizontal runners which were made a part of the supporting scaffold. The collection trough was positioned underneath the tower to collect a given percentage of the tower throughput. The actual flow through the entire filter system was controlled by positioning the effluent collection trough.

The use of amphoteric metal, such as copper or zinc, was held to a minimum. In particular, copper tubing was avoided. Brass valves were acceptable.

Because of the height of the tower and the need for easy access, the tower was erected next to a building where a ladder was already in place. It was necessary to observe the operation of the tower from the top. Maintenance of the 1/2 inch solenoid valve also required access to the top.

APPENDIX D

MOBILE ACTIVATED SLUDGE PILOT PLANT (INSTALLATION)

I. Requirements

A. A level area was selected (10' x 15') close to the sample point.

B. Electricity used was 440-480 volt 3 phase alternating current (220 Volt A.C. can be used). Since high voltage plugs vary from place to place, a male plug was needed for the unit. The unit was equipped with 50 feet of power cord.

C. Hose - 3/4" nylon reinforced garden hose was used. Three hoses from the pilot plant to the point being sampled were required.

II. Set Up

A. The unit was designed to run by gravity flow. This was the reason for selecting a level site.

B. The available current should be determined. If it is 220 volt A.C. the taps must be changed on the transformer, pump motor, and blower motor. The taps are located where the power line connects and are marked. The heater elements in the junction box inside the panel must also be changed. The elements are stored in the panel and are marked for required changes. If 440-480 volt A.C. current is available, no changes are needed.

C. Hose connections:

1. A hose was connected from the pump intake to the sample source.

2. A hose was connected from the manual valve on the pump outlet to the sample source.

3. A hose marked A-settler connected the solenoid on the pump outlet to the interior tap on the primary settler marked A on diagram.

4. A hose marked Aerator connected the outside tap on the primary settler to the tap on the solenoid valves on the aerator (Tank B).

5. A hose marked B-Settler connected the ports on the side of the aerator to the solenoid valves on the secondary settler Vessel C.

6. A line marked Air Lift connected the tap by the air flowmeter to the inside tap on the secondary settler. Care was taken to loop this line over the top of the pipe by the settler so that the line was higher than the settler.

7. A third line connected the untreated waste to the unit and was attached to the outside tap on secondary settler.

8. An air lift line was attached to the air input port in the secondary settler and to the air lift lodged at the base of the settler.

9. Weirs on the two settlers were attached to the outside ports.

III. Start Up

A. The water pump was primed by means of the tap on the top of the pump housing.

B. All switches on the master control panel were off.

C. The main power switch (on the rear of the panel) was switched ON.

D. The manual valve on the pump outlet was fully opened.

E. The main power was switched to the ON position.

F. The main pump switch was turned to the ON position.

G. When the pump drew water from the untreated waste and returned it to the same point, the system was ready to start.

H. The level control switch was turned to ON. This opened the solenoid and allowed liquid to flow into the primary settler.

I. The interval timer on the primary settler side of the panel was positioned to 50% and turned on.

J. When the aeration vessel was approximately 1/2 full, the blower switch was turned to ON. The aeration rate in each chamber was adjusted by means of the ball valves on the air lines such that they appeared to be slightly boiling.

K. Sludge was added at this point.

L. When the aeration vessel was filled to the level of the top exit ports, the interval timer on the secondary settler side of the unit was positioned at 50% and turned on.

M. If no flow occured from the ports through the secondary settler solenoids, the end of the hose from the solenoids was removed and lowered until a flow resulted. Then the hose was reconnected.

N. When the secondary settler was 3/4 full, the ball valve controlling the air lift was cracked slightly. After small adjustments, the air and water started flowing up the air lift tube. This tube was placed over the first chamber of the aeration vessel to return the sludge.

APPENDIX E

MOBILE ACTIVATED SLUDGE PILOT PLANT (OPERATION)

The pilot plant was equipped with a feed tank capable of holding a two day supply. The tank was filled with combined coke plant effluent waste, exclusive of the coal washery effluent. Alkaline dumps resulting from caustic washings of benzol products were also excluded from the feed to the pilot unit.

When the feed tank was filled with waste, it was immediately checked for cyanide, phenol, and ammonia concentrations. It was desirable to keep the ammonia concentrations at levels below 1500 ppm. The feed batch was also checked for pH, phosphates and COD. If the COD: PO_4 ratio was less than 100:3, phosphate was added until this ratio had been established. Tri-sodium phosphate was used for this purpose. The pH was in the range of 6.5-8.0.

When the coke plant waste was treated by the activated sludge process, both phenol and cyanide were consumed by the micro-organisms in the activated sludge. It was assumed, however, that the loading data obtained as indicated above was based primarily on phenol removal. COD removal would also be a suitable criterion.

The batch studies referred to above established the ratio of phenol (or COD) concentration to MLVSS which gave the maximum rate of phenol and COD removal. The pilot plant was started and operated at this loading.

Aeration rate was adjusted so as to maintain a D.O. level in the aeration chamber of 4.0 mg/l.

Variation of flow rates actually represented a variation in detention time in the activated sludge system. The MLVSS in the aeration chamber was kept constant while the flow rates were varied. Adjustment of MLVSS was accomplished by adjusting sludge returned from the final clarifier to the aeration basin. The determination of rates of sludge formation were essential when the system was operating under a loading condition. A material balance was obtained as follows:

Starting with no sludge in the final clarifier, it was possible to account for all sludge generated by allowing the sludge to build up in the final clarifier for a given period of time.

Total quantities of sludge generated were also determined by collecting all waste sludge.

APPENDIX F

OPERATING LOG SHEETS OF THE ACTIVATED SLUDGE PILOT PLANT

Project No 12010 DWF

APPENDIX F

Activated Sludge Pilot Plant

	Average				D.O.	COD		5 Day BOD		Phenol		Cyanide		Ammonia		Phosph
	Temp °F		pH		ppm	ppm		ppm		ppm C_6H_5OH		ppm CN		Nitrogen ppm		
	IN	OUT	IN	OUT	Aer. Basin	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	Aer
4-1-69 Single Aer. Basin in Service				6.0						454	8.0	74	25	170	182	
4-2-69 - Added Small sludge From batch				7.4										160		
4-3-69				7.4												
4-4-69				7.4												
4-5-69 - added Small sludge From batch				7.4												
4-6-69																
4-7-69 - added sludge From batch Tank																
4-8-69																
4-9-69																
4-10-69						159										[illegible]
4-11-69				7.2												
4-12-69				7.1												
4-13-69	51	54														
4-14-69 - Put all (3) Aer. basins in Service	56	59		7.0												
4-15-69	51.5	55		7.0												
4-16-69	40.5	45		6.9												
4-17-69	44	49		7.0						421	2.8					
4-18-69 - Pumped Sludge From Sec. Set. To Aer. Basins	63.4	58		7.0												
4-19-69	61	58		7.0												
4-20-69	70	67		7.0												
4-21-69	65	64		7.3						17.5	.25	3.6	2.2	93.5	59.5	
4-22-69	68	65		7.3	6.2					16.5	.57	2.0	1.7	95.0	55.3	
4-23-69 Primary Carbon mixed with EFF.	64	61		7.5						8.5	.10	3.0	1.7	89.3	34.8	
4-24-69	69	67		7.3	6.0					28.4	7.9	7.2	2.04	225	221	
4-25-69 increased Air to Basins	56	63.5		7.9												
4-26-69	48	51														
4-27-69	51	57.4														
4-28-69	56	54		6.9	8.0					28.4	.30	6.46	.44	238	221	
4-29-69	65	61.6		7.3	7.6					31.6	.80	10.2	1.6	229	221	
4-30-69	71	68		7.6	7.1					33.3	.30	5.08	1.15	229	229	
5-1-69	71	69		7.7	6.4					9.6	.24	6.06	.87	225	221	
5-2-69	70.5	67		7.0	6.8											
5-3-69 increased Flow	64	64														
5-4-69		51														
5-5-69		50		6.8												
5-6-69		48		6.8	5.0											
5-7-69 - added sludge From Batch Tank	50	58		7.8	6.0											
5-8-69 "	71	57		7.8	6.2											
5-9-69	64	63		7.8	6.1											

APPENDIX F

D.O. = Dissolved Oxygen
VSS - Volatile Susp Solids
SVI - Sludge Volume Index

tivated Sludge Pilot Plant

COD ppm IN	COD ppm OUT	5 Day BOD ppm IN	5 Day BOD ppm OUT	Phenol ppm C_6H_5OH IN	Phenol ppm C_6H_5OH OUT	Cyanide ppm CN IN	Cyanide ppm CN OUT	Ammonia Nitrogen ppm IN	Ammonia Nitrogen ppm OUT	Phosphate ppm Aer Basin	Mixed Liquor Susp Solids ppm Aer Basin	VSS ppm Aer Basin	Total Solids ppm Aer Basin	Flow Inlet mL/min.	Flow Sludge Return mL/min.	Sludge Volume mL/L
				954	8.0	74	25	170	172		178			25		<1
								160			169			25		<1
																1.3
																2.5
																3.0
																11.0
																9.0
																29.0
														25		35.0
159										32	960			25		30
											860			25		30
											890			25		30
														25		25
														75		35
											458			120		18
														120		10
				428	2.8									120		10
														120		
											517			120		18
														120		
				17.5	.25	3.6	2.2	93.5	59.5		510			160		16
				16.5	.57	2.0	1.7	95.0	55.3		448			160		16
				8.5	.10	3.0	1.7	89.3	84.8					160		10
				28.4	7.9	7.2	2.04	225	221		492			200		14
											120			200		10
				28.4	.30	6.16	.44	238	221		139			200	1,000	8
				31.6	.80	10.2	1.6	229	221		910			200		10
				33.3	.80	5.08	1.5	229	229		575			200		15
				9.6	.34	6.06	.89	225	221		431			200	1,000	9
														200		10
														275		
														275	1,000	10
											597	312		275		
											1030	192		275	1,000	3
											565	242		275		6
											1,013	428		275		10
											654	283		275		50

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PROJECT NO 12010 DNF

APPENDIX F

Actuated Sludge Pilot Plant

	Average				D.O.	COD		5 Day BOD		Phenol		Cyanide		Ammonia	
	Temp °F		pH		ppm	ppm		ppm		ppm C₆H₅OH		ppm CN		N Kjeldahl ppm	
	IN	OUT	IN	OUT	Aer. Basin	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT
5-10-69	65	58													
5-11-69	56	57													
5-12-69	59	61													
5-13-69	61	63				1442	58								
5-14-69	66	67													
5-15-69	66	60													
5-16-69	49	49		7.7	7.1	717	587								
5-17-69	55	58													
5-18-69	85	66		7.5											
5-19-69	93	69		7.8	6.0	981	812			4.8	.15	3.72	.80	78.1	78
5-20-69	90	60		7.5	6.8	767	579			2.7	.10	2.0	.30	105	77
5-21-69	84	60		7.0	6.3	590	209			30	3.6	5.97	1.31	92	77
5-22-69	83	55		7.4	7.6	900	221								
5-23-69	79	58		7.1											
5-24-69 Reduced aeration + sludge return	86	62													
5-25-69	75	68													
5-26-69	82	75		6.7	6.0	307	146			23.8	.21	5.7	.56	30	25.5
5-27-69	84	73		8.0	6.0	607	227								
5-28-69	81	73		7.75	5.8	857	381								
5-29-69	78	71		7.65		596	359			63.5	1.7	30.7	3.45	232	218
5-30-69	87	73													
5-31-69	80	62													
6-1-69	62	58													
6-2-69	73	65		7.4	5.7										
6-3-69	72	68	8.5	8.0	5.8										
6-4-69	85	75	8.5	8.0	5.8		578								
6-5-69	84	79	7.9	7.8	4.5	1372	882								
6-6-69	93	76	7.8	7.9	2.5										
6-7-69	72	66	7.0	7.5		394	279	2650	164						
6-8-69 Air line plugged	68	61	7.3	7.35	4.0										
6-9-69 "	74	66	8.15	7.5	5.7	393	228	7950	65	8.2	.17	11.6	3.98	218	212.5
6-10-69	72	69													
6-11-69	57	54	7.5	7.55	4.9	323	227								
6-12-69 Power Failure. Feeding system manually	71	59	7.8	7.7	5.9	195	97	430	66						
6-13-69	63	54			5.0										
6-14-69 "	65	53													
6-15-69 "	58	54													
6-16-69 "	54	57	8.4	8.1	4.7	506	106								
6-17-69 "	63	57	8.3	7.6	4.7	594	205								
6-18-69 "	68	65	7.8	8.0	6.8	374	485								

APPENDIX F

D.O. : Dissolved Oxygen
VSS - Volatile Susp. Solids
SVI : Sludge Volume Index

ated Sludge Pilot Plant

COD		5 Day BOD		Phenol		Cyanide		Ammonia		Phosphate	Mixed Liquor Susp. Solids	VSS	Total Solids	Flow		Sludge
ppm		ppm		ppm C6H5OH		ppm CN		N.Trogen ppm		ppm	ppm	ppm	ppm	Inlet	Sludge Return	Volume
	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	AER. BASIN	AER. BASIN	AER BASIN	AER BASIN	ML/MIN	ML/MIN	ML/L
														50	1,000	
														50		
														275	1,000	
[illegible]	58													275		
														275		
[illegible]	587										700			275	1,000	3
														275		
											824	180		275		7
[illegible]	812			4.8	.15	3.72	.80	78.1	78		1,128	239		275	1,000	5
[illegible]	579			2.7	.10	2.0	.30	105	77		509	189		275		2.5
[illegible]	209			30	3.6	5.97	1.31	82	77		333	116		300		2.5
[illegible]	221									90	479	296		250	1,000	2.5
											1,005	87		250		2.5
														250		3
														275		6
[illegible]	146			23.8	.21	5.7	.56	30	25.5	42	66	4		300	150	1
[illegible]	227									50	180	85		250		1
[illegible]	381									65	240	11		275		3
[illegible]	359			625	1.7	30.7	3.45	232	213	75	206	104		250		2
														180	150	6
														175		6
														250		10
											633	326		250		9
																12
	538									90	774	441		250	150	21
[illegible]	882									90	795	338		200	150	30
										90	840	324		225	150	35
[illegible]	279	2650	164								891	205		275	150	50
											743	219		205	-	30
[illegible]	228	2450	65	82	.17	11.6	3.88	218	212.5		176	59		350	-	1
														225	150	
[illegible]	237									90†	576	83		250	150	18
[illegible]	97	450	66								488	166				15
																20
																10
[illegible]	106													150	400	10
[illegible]	208										489	116		400	250	12
[illegible]	485										415	129	2,754		275	10

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PROJECT NO. 12010 DNF

APPENDIX F

Activated Sludge Pilot Plant

			Average				D.O.	COD		5 Day BOD		Phenol		Cyanide		Ammonia		Phosphate
			Temp. °F		pH		ppm	ppm		ppm		ppm C_6H_5OH		ppm CN^-		Nitrogen ppm		ppm
			In	Out	In	Out	Aer. Basin	In	Out	In	Out	In	Out	In	Out	In	Out	Aer. Basin
6-19-69	Power System Failure Feeding		73	70	7.2	7.5	4.6	464	304	240	150							
6-23-69	System Manually				6.3	6.5	6.6	482	65									
6-24-69	"				7.85	6.4	4.4	570	134			21	.5					
6-25-69	"				7.75	6.6						40	.35					
6-26-69	Normal Operation				7.2	6.8	4.2	689	151					10.3	.8			
6-27-69	" Solenoid Sticking		79	76	7.5	6.9	6.2	1,330	526					7.2	8.0			
6-30-69	Normal Operation		81	78	7.5	7.2	3.6	1,369	371			29	.15					
7-1-69	" Switched Sludge Return through Solenoid Valve		79	76	7.8	7.6	<1.0	334	76				6.9					
7-2-69	"		87	79	7.5	7.1	1.9	358	313			35	.17					
7-3-69	" Found 1st Aer. Basin outlet plugged & no. air		88	83	7.0	6.7	3.3	498	228	230	150	29.1	1.7	8.1	1.17	289	258	
7-4-69	"		78	69	6.85	6.9	2.5	1										
7-5-69	"		78	72														
7-6-69	"		80	75														
7-7-69	"		83	79	7.5	7.25	2.7	381	273			13.5	.3	2.1	.24			
7-8-69	"		82	78	7.35	7.15	4.6	547	466			33.3	1.3	4.2	.40			
7-9-69	"		78	74	7.0	7.8	2.5	732	366									
7-10-69	"		80	76	6.6	6.5	2.0	602	372									
7-11-69	"		90	81	6.2	6.6	1.6	422	318	150	42							
7-12-69	"		85	79														
7-13-69	"		89	82														
7-14-69	"	9:00 AM	85	76			2.6											
(Out cycle storage TK.)		10:00 AM	94	78														
		11:00 AM	98	83														
		12:00 PM	102	86	7.9	7.4												
		1:00 PM	104	90	7.9	7.4												
		2:00 PM	106	94	7.9	7.4		425	203	170	40	375	.09	5.1	1.0	255	170	
		4:00 PM	106	99	7.9	7.4												
		5:00 PM	101	96	7.9	7.4												
		6:00 PM	101	95	8.0	7.5												
		7:00 PM	96	93	8.0	7.5												
		8:00 PM	94	90	8.0	7.5												
		9:00 PM	90	86	8.0	7.5												
		10:00 PM	86	81	8.0	7.5		418	189	190	45	49.5	.16	5.2	.78	323	204	
7-15		12:00 AM	85	78	7.85	7.55												
		1:00 AM	85	76	7.85	7.55												
		2:00 AM	83	74	7.8	7.65												
		3:00 AM	82	73	7.75	7.55												
		4:00 AM	80	72	7.6	7.5												
		5:00 AM	79	70	7.6	7.65												
		6:00 AM	77	68	7.5	7.55		364	187			355	.10	2.83	.73	357	221	

APPENDIX F

Activated Sludge Pilot Plant

COD ppm IN	COD ppm OUT	5 Day BOD ppm IN	5 Day BOD ppm OUT	Phenol ppm C_6H_5OH IN	Phenol ppm C_6H_5OH OUT	Cyanide ppm CN IN	Cyanide ppm CN OUT	Ammonia Nitrogen ppm IN	Ammonia Nitrogen ppm OUT	Phosphate ppm Aer. Basin	Mixed Liquor Susp. Solids ppm Aer. Basin	VSS ppm Aer. Basin	Total Solids ppm Aer. Basin	Flow Inlet mL/min	Flow Sludge Return mL/min	Sludge Volume ml/l	
[illegible]	304	240	150								194	41	2,074		150	1	
[illegible]	65										171	60	1,736		150	1	
[illegible]	184			21	.5						551	188	2,177	350	175	1	
				40	.35						980	220	2,219		200	18	
[illegible]	151					18.3	.8				250	90	1,236	250	150	5	
[illegible]	526					7.2	8.0				890	298	1,080	-	200	7	
[illegible]	371			29	.15						476	119	1,445	370	650	1	
[illegible]	76				6.9						185		949	375	200	1	
[illegible]	303			95	.17												
[illegible]	289	230	130	20.1	1.7	8.1	1.17	299	258		723	210	1,835	250	250	15	
											754	280	1,654	260	250	18	
														260	310	35	
														260	290	35	
[illegible]	278			135	.8	2.1	.34				1,243	277	2,260	260	325	45	
[illegible]	466			33.5	1.3	4.2	.40				935	194	2,049	250	275	40	
[illegible]	866										1,266	303	2,654	250	310	30	
[illegible]	372										1,970	703	2,836	215	230	37	
[illegible]	318	150	42								1,965	425	2,839	250	270	45	
														285	305	45	
														285	305	40	
											1,506	335	2,161	228	372	38	
														280	372	40	
														200	362	37	
														225	305	39	
														350	335	39	
[illegible]	207	170	40	375	.09	6.1	1.0	265	170					285	382	39	(Comp.)
														280	330	38	
														290	305	39	
														290	320	38	
														285	305	39	
														290	320	39	
														255	300	39	
[illegible]	189	190	45	485	.16	5.2	.39	223	204		1,325	232		285	300	39	(Comp.)
														280	290	39	
														280	295	40	
														280	295	39	
														280	300	40	
														280	305	39	
														280	305	38	
[illegible]	187			355	.10	387	.72	357	221		1,408	292		290	300		(Comp.)

B

Project No 12010 DNF

APPENDIX F

Activated Sludge Pilot Plant

		Average				D.O	COD		5 Day BOD		Phenol		Cyanide		Ammonia		phosphate
		Temp °F		pH		ppm	ppm		ppm		ppm C_6H_5OH		ppm CN		Nitrogen ppm		ppm
		in	out	in	out	Aer. Basin	in	out	in	out	in	out	in	out	in	out	Aer. Basin
7-15-69 Normal operation	8 am	78	69	7.35	7.50	1.8											
	9:00 AM	80	72	7.20	7.5												
	10:00 AM	85	78	7.3	7.5												
	11:00 AM	91	84	7.5	7.5												
	12:00 PM	94	87	7.3	7.5												
	1:00 PM	96	92	7.2	7.5												
	2:00 PM	98	94	7.2	7.5		290	251			10	.10	2.5	.87	340	301	
	4:00 PM	97	96	7.3	7.5												
	5:00 PM	98	94	7.4	7.5												
	6:00 PM	96	93	7.4	7.5												
	7:00 PM	96	92	7.5	7.55												
	8:00 PM	91	89	7.5	7.6												
	9:00 PM	90	87	7.5	7.6												
	10:00 PM	86	84	7.6	7.65		153	112			5.5	.10	4.05	1.13	340	306	
	12:00 AM	85	80	7.6	7.6												
	1:00 AM	84	77	7.7	7.6												
	2:00 AM	82	75	7.7	7.6												
	3:00 AM	81	74	7.7	7.6												
	4:00 AM	81	74	7.7	7.6												
	5:00 AM	79	73	7.7	7.6												
	6:00 AM	79	72	7.8	7.6		161	83			1.5	Trace	.94	.52	340	238	
7-16-69 "	8:00 AM	77	71	7.8	7.7	8.0											
	9:00 AM	80	74	7.8	7.65												
	10:00 AM	85	79	7.8	7.6												
	11:00 AM	89	83	7.8	7.6												
	12:00 PM	93	86	7.75	7.6												
	1:00 PM	96	89	7.7	7.6												
(End of Storage Tank Cycle)	2:00 PM	99	93	7.7	7.55						1.5	.17	.80	.74			
7-17-69 "		85	78	7.5	7.6	1.5					4.5	Trace	2.46	.27	340	306	
7-18-69 "	8 am	81	76	7.9	7.7	8.3											
	3 pm	83	79	8.0	7.9						.38	Trace	1.35	.51			
	11 pm	80	76	8.0	7.8						.38	Trace	2.0	.82	153	136	
7-19-69 "		80	72														
7-20-69 "		76	69														
7-21-69 "	8 am	87	74	8.9	7.5	31											
	11 pm																
7-22-69 "		83	76	8.1	7.5	1.9											
7-23-69 "		89	82	7.85	7.6	3.7											
7-24-69 "		81	76	7.83	7.62	3.1					21.5	Trace	10.4	.75	68	68	
7-25-69 " (no feed)		79	76			3.6											

A

APPENDIX F

ActivaTed Sludge PiloT PlanT

	5 Day BOD		Phenol		Cyanide		Ammonia		phosphate	Mixed liquor Susp. Solids	VSS	Total Solids	Flow		
	ppm		ppm C_6H_5OH		ppm CN		NITROGEN ppm		ppm	ppm	ppm	ppm	INLET ml/min	Sludge Return ml/min	Sludge Volume ml/l
OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	AER. BASIN	AER. BASIN	AER. BASIN	AER. BASIN			
													285	285	39
													285	345	39
													290	300	38
													290	325	38
													280	382	39
													285	370	38
251			10	.10	2.5	.87	340	301		1,152	242		290	285	40 (comp)
													290	370	38
													285	330	39
													285	370	38
													285	320	38
													290	295	39
													280	310	38
112			5.5	.10	4.05	1.13	340	306		1,447	346		280	310	39 (comp)
													290	355	38
													290	305	38
													280	285	37
													280	300	37
													280	285	36
													280	300	
83			1.5	TRACE	.94	.52	340	238					290	290	(comp)
													280	280	36
													285	350	33
													290	305	30
													295	305	31
													260	300	32
													285	310	30
			1.5	.17	.80	.74				1,074	419		310	370	27 (comp.)
			4.5	TRACE	2.46	.27	340	306		1,386	327	2,631	230	285	30
										1,161	246	1,918	260	275	30
			.38	Trace	1.35	.51									25
			.38	trace	2.0	.82	153	136							23
													290	335	30
													295	320	30
										882	191	1,186			24
										834	220	1,028	280	330	24
										867	189	1,000	310	280	24
			21.5	TRACE	10.4	.75	68	68		626	140	4,279	290	305	24
										842	240			250	12

B

Project No. 12010 DNF

APPENDIX F

Activated Sludge Pilot Plant

			AVERAGE				D.O.	COD		5 Day BOD		Phenol		Cyanide		Ammonia		Phosphate		
			Temp °F		pH		ppm	ppm		ppm		ppm C_6H_5OH		ppm CN		Nitrogen ppm		ppm		
			IN	OUT	IN	OUT	Aer. Basin	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	
7-28-69	(Feed pump Down)		87	78	8.0	7.25	5.0													
7-29-69	Normal Operation		88	82	7.8	7.8	2.1													
7-30-69	"	7^{am}	86	79	7.2	7.4	1.3					6.8	.23	7.2	.44					
	"	3^{pm}												6.4	.63					
7-31-69	"	7^{am}	79	77	7.0	7.05	2.2							4.0	.37					
	"	3^{pm}												2.6	.33					
	"	11^{pm}												1.45	.35					
8-1-69	"		86	81	7.1	7.4	3.0							2.15	.30	119	102			
8-2-69	"		86	79																
8-3-69	"		84	79																
8-4-69	"	7^{am}	83	69	7.25	7.35	1.5									271	171			
	"	3^{30} pm	90	92												267	155			Duplicates
																267	153			
	"	5^{30} pm	88	89												272	158			Duplicates
																272	156			
	"	7^{30} pm	88	85												270	170			Duplicates
																270	170			
	"	9^{00} pm	87	92												270	178			Duplicates
																270	177			
8-5-69	"	7^{00} am	70	79	7.0	7.2	.9	488	310					5.66	.39	358	326			
																420	384			40 ppm added
		11^{00} am	90	83												388	342			Duplicates
																391	340			
8-6-69	B.O. Solenoid	8^{00} am	75	85	7.23	7.3	2.8	775	550							306	268			
																350	309			40 ppm added
	Normal Operation	11^{00} am	70	83												306	269			Duplicates
																309	270			
	"	1^{00} pm	68	94												309	284			
																347	323			40 ppm added
	"	3^{00} pm	65	99												309	289			
																348	330			40 ppm added
8-7-69	"	7^{00} am	63	70	7.3	7.5	3.4							2.0	.29					
	"	4^{00} pm	100	91												313	287			Duplicates
																309	289			
	"	6^{00} pm	95	89												311	287			
																348	328			40 ppm added
	"	8^{00} pm	97	83												337	294			Duplicates
																339	292			

APPENDIX F

ActivATEd Sludge PilOT PlANT

COD		5 Day BOD		PHENOL		cyanide		Ammonia		Phosphate		Mixed liquor Susp. Solids	VSS	Total Solids	Flow		
ppm		ppm		ppm C_6H_5OH		ppm CN		Nitrogen ppm		ppm		ppm	ppm	ppm	Inlet	Sludge Return	Sludge Volume
IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	Aer. Basin	Aer. Basin	Aer. Basin	ml/min.	ml/min.	ml/l
												362	162	600		350	12
												415	197	836	320	365	6
				6.8	.23	7.2	.44					549	289	2,233	352	300	5
						6.4	.63								352		
						4.0	.37					481	315	1,214	275	315	9
						2.6	.33										
						1.45	.35										
						2.15	.30	119	102			618	392		180	355	12
															394	305	21
															166	325	28
								271	171			957	574	1,401	340	300	18
								267	155	Duplicates							
								267	153								
								272	158	Duplicates							
								272	156								
								270	170	Duplicates							
								270	170								
								270	178	Duplicates							
								270	177								
88	310					5.66	.39	388	326			1,234	772	2,226	415	365	18
								420	384	40 ppm added							
								388	342	Duplicates							
								391	340								
773	550							306	268								
								350	309	40 ppm added		1,104	670	2,129	280	350	20
								306	269	Duplicates							
								309	270								
								309	284								
								347	323	40 ppm added							
								309	289								
								348	330	40 ppm added							
						2.0	.29								280	230	20
								313	287	Duplicates							
								309	289								
								311	287								
								348	328	40 ppm added							
								337	294	Duplicates							
								339	292								

PROJECT NO. 12010 DNF

APPENDIX F

ACTIVATED Sludge Pilot Plant

		AVERAGE				D.O	COD		5 DAY BOD		PHENOL		Cyanide		Ammonia		Phosphate
		TEMP °F		pH		ppm	ppm		ppm		ppm C_6H_5OH		ppm CN		NITROGEN ppm		ppm
		IN	OUT	IN	OUT	AER. BASIN	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN
8-8-69	NORMAL OPERATION	92	85	7.5	7.1												
8-9-69	"	81	75	7.5	7.1												
8-10-69	"	82	76	7.3	7.0												
8-11-69	"	86	79	7.3	6.9												
8-12-69	"	89	80	7.25	7.0		372	327									
8-13-69	" 4:00 PM	103	94	7.3	7.1		411	352									
	" 3:00 AM	85	76	7.3	7.1		405	286									
8-14-69	"	79	71	7.7	7.4												
8-15-69	"	83	76	7.7	7.4												
8-16-69	"	89	81	7.5	7.4												
8-17-69	"																
8-18-69	" 7AM - 3PM	99	91	7.0	7.25	3.5	777	299			36	.1	13.6	.27	677	243	
	" 3PM - 11PM	87	83	7.2	7.0	3.5	790	410			36.5	.1	8.6	2.05	612	357	
	" 11PM - 7AM	85	80	6.8	6.9	2.5	911	385			34.5	.1	6.5	3.0	629	489	101
8-19-69	" 7AM - 3PM	85	80	7.1	7.15	2.3		272				.1		.95		510	
	" 3PM - 11PM	90	85	7.1	7.15	2.3		239				.1		.2		569	
	" 11PM - 7AM	78	73	7.1	7.15	2.5	536	321			38	TRACE	2.8	.23			99
8-20-69	" 7AM - 3PM	88	82	7.9	6.9	2.8	771	252			36	TRACE	19.2		442		63
	" 3PM - 11PM	92	87		7.0	2.8		471				TRACE		1.96		213	
	" 11PM - 7AM	80	73	7.7	7.2	3.5	511	325			30	TRACE	2.8	.5	272	221	
8-21-69	" 7AM - 3PM	89	84	7.7	7.2	2.5		318				TRACE		1.8		238	
	" 3PM - 11PM	97	88	7.7	7.15	2.5						TRACE		2.4		204	
	" 11PM - 7AM	85	76	7.7	7.2	3.0	457	326			7.3	TRACE	2	.2	187	191	70
8-22-69	" 4:00 PM	93	80	7.7	7.5		430	329									
8-23-69	"	82	67														
8-24-69	"	87	69														
8-25-69	" 7AM - 3PM	89	74	8.35	7.4	4.0	852	520			42	1.2	13.2	.3	616	485	120
	" 3PM - 11PM	85	80	8.0	7.5	3.0		504				.6		.7		604	
	" 11PM - 7AM	77	66	8.0	7.7	2.5	656	336			38	.2			612	598	
8-26-69	"	91	68														
8-27-69	" 7AM - 3PM	89	79	8.4	7.4	3.5	460	406			24	.11	6.44	1.5	340	306	325
	" 3PM - 11PM	88	86	8.4	7.8	2.5		381				.12		.54		230	
	" 11PM - 7AM	85	70	8.4	7.8	2.3	369	287			1.0	.08	2.72	.67	221	212	
8-28-69	"	83	77														
8-29-69	"	73	77	7.9	7.75						1.0	.083	2.72	.71	110	127	
8-30-69	"	72	70	7.8	7.6												
8-31-69	"	79	67	7.3	7.1												
9-1-69	"	119	65	7.3	7.15		189										

APPENDIX F

Activated Sludge Pilot Plant

COD ppm OUT	5 Day BOD ppm IN	5 Day BOD ppm OUT	Phenol ppm C6H5OH IN	Phenol ppm C6H5OH OUT	Cyanide ppm CN IN	Cyanide ppm CN OUT	Ammonia Nitrogen ppm IN	Ammonia Nitrogen ppm OUT	Phosphate ppm IN	Phosphate ppm OUT	Mixed Liquor Susp. Solids ppm Aer. Basin	VSS ppm Aer. Basin	Total Solids ppm Aer. Basin	Flow Inlet ml/min	Flow Sludge Return ml/min	Sludge Volume ml/l
														280	310	
														340	365	
														330	370	
														330	320	
327														280	335	
352											697	392		300	320	
286											704	410		305	310	
														300	310	
														360	340	
														380	340	
299			36	.1	13.6	.27	687	243			1,385	729		331	350	>40 comp
410			36.5	.1	8.6	2.05	612	357						331	350	" "
385			34.5	.1	6.5	3.0	629	489	101	104	1,307	733		331	350	" "
272				.1		.95		510			1,838	979		331	350	" "
239				.1		.2		569			1,913	1,003		331	350	" "
321			38	Trace	2.8	.23			99	103	2,002	1,084		331	350	" "
252			36	Trace	19.2		442		63		2,126	1,117		485	350	" "
471				Trace		1.96		213		28	2,027	1,120		485	350	" "
325			30	Trace	2.8	.5	272	221			2,210	1,193		485	350	" "
318				Trace		1.8		238			2,417	1,275		485	350	" "
				Trace		2.4		204			2,399	1,300		485	350	" "
326			7.3	Trace	2	.2	187	191	70	55	2,344	1,234		485	350	" "
329											2,621	1,242		350	550	
520			42	1.2	13.2	.3	616	485	120		1,478	803		490	350	comp
504				.6		.7		604			1,566	872		490	350	"
336			38	.2			612	598			1,679	931		490	350	"
406			24	.11	6.44	1.5	340	306	325		2,292	1,210		939	350	comp.
381				.12		.54		230			2,793	1,515		939	350	"
287			1.0	.08	2.72	.67	221	212			2,007	1,014		939	350	"
			1.0	.083	2.72	.71	110	127						380	350	
														350	310	
														350	350	
											981	518		320	345	

B

PROJECT NO. 12010 DNF

APPENDIX F

ACTIVATED SLUDGE PILOT PLANT

		AVERAGE				D.O.	COD		5 DAY BOD		PHENOL		CYANIDE		AMMONIA		PHOSPHAT	
		TEMP °F		pH		ppm	ppm		ppm		ppm C2H5OH		ppm CN		NITROGEN ppm		ppm	
		IN	OUT	IN	OUT	AER. BASIN	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT
9.2-69	NORMAL OPERATION 7 AM	73	75	8.0			595				12.5		15		161			
	" Comp. 1:30-3:30 PM	102	80		7.6			338				1.58		3.5		145		
	" Comp. 4:00 4:45 PM	101	95		7.6			434				3.17		7.1		145		
9.3-69	"	78	76															
9.4-69	" 7 AM	74	64	7.5			601						15.8		225			
	" Comp. 10:30-11:30 AM	74	88		7.55			352			52	TRACE		.9		119		
Sludge Return Plugged	" Comp. 1:00 2:00 PM	80	100		7.6			358				13.6		2.75		229		
	" Comp. 2:15-3:15 PM	83	104		7.55			362				22.8		3.10		229		
9.5-69	" 7 AM	69	81	7.7			452				39		3.2		706			
	" Comp 9:00-10:00 AM	73	75		7.5			266				6.38		1.1		637		
	" Comp 10:30-11:30 AM	77	72		7.5			339				4.59		1.7		659		
	" Comp 12:00-1:00 PM	87	69		7.5		364	323			30	1.6	1.8	1.35	646	646		

APPENDIX F

Activated Sludge Pilot Plant

COD		5 Day BOD		Phenol		Cyanide		Ammonia		Phosphate		Mixed Liquor Susp. Solids	VSS	Total Solids	Flow		
ppm		ppm		ppm C_6H_5OH		ppm CN		Nitrogen ppm		ppm		ppm	ppm	ppm	Inlet	Sludge Return	Sludge Volume
	Out	In	Out	In	Out	In	Out	In	Out	In	Out	Aer. Basin	Aer. Basin	Aer. Basin	ml/min	ml/min	ml/l
				12.5		15		161				931	518		1650	350	Comp
	338				1.58		3.5		145						1650	350	"
	434				3.17		7.1		145			1,185	555		1650	350	"
						15.8		225							1060	350	
	352			52	TRACE		.9		119						1060	300	Comp
	358				13.6		2.75		229						1060	–	"
	362				22.8		3.10		229						1060	–	"
				39		3.2		706				1,574	817		1210	350	
	266				6.38		1.1		637						1210	350	Comp
	339				4.59		1.7		659						1210	350	"
	323			30	1.6	1.8	1.35	646	646			1,484	718		1210	350	"

B

APPENDIX G

Activated Carbon Specifications

Physical Properties	Filtrasorb 300	Filtrasorb 400
Total Surface Area (N_2 Bet Method), m^2g	950-1050	1000-1100
Apparent Density (Bulk Density, dense packing), g/cc	0.48	0.44
Lb./Ft^3	30.0	27.5
Particle Density (Hg Displacement), g/cc	0.75	0.7
Real Density (He Displacement), g/cc	2.1	2.1
Pore Volume (Within Particle), cc/g	0.85	0.94
Voids in Dense Packed Column %	36	38
Specific Heat at 100° C	0.25	0.25
Specifications		
Mesh Size, U.S. Sieve Series	8 x 30	12 x 40
Iodine Number, Minimum	900	1000
Molasses Number, Minimum	200	230
Ash, Maximum %	8.0	8.5
Moisture as packed, Maximum %	2.0	2.0
Abrasion Number, Minimum	70	75
Lot Number		9C3-F
Date		9/1/69

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APPENDIX H

Analytical Methods

1a) Ammonia Nitrogen - FWPCA Interim Methods for Chemical Analysis of Surface Waters, September 1968. Nitrogen-Ammonia by distillation.

1b) Nitrogen-Kjeldahl - "Standard Methods" 12th Ed., 1965 pp. 209-211.

2) Chemical Oxygen Demand - "Standard Methods" 12th Ed., 1965 pp. 510-514.

3) Chloride - "Standard Methods" 10th Ed., 1955 p. 60, Section B.

4) Color - Hach Chemical Company, Water and Wastewater Analysis Procedures, August 1967.

5) Cyanide - "Standard Methods" 12th Ed., 1965 Methods A&B pp. 453-457. (Test modified by use of tartaric acid in place of sulfuric acid to avoid sulfide interference)

6) Dissolved Oxygen - Yellow Springs Dissolved Oxygen Meter and "Standard Methods" 12th Ed., 1965 Modified Winkler Method pp. 406-410.

7) Ether Extractables - "Standard Methods" 12th Ed., 1965 pp. 211-213. (Carbon tetrachloride was used in addition to petroleum ether)

8) Nitrates - "Standard Methods" 12th Ed., 1965 Method B pp. 198-200.

9) pH - "Standard Methods" 12th Ed., 1965 pp. 226-228.

10) Phenol - "Standard Methods" 12th Ed., 1965 Methods A&C pp. 517-523.

11) Phosphate - Ortho-Phosphate - "Standard Methods" 12th Ed., 1965 Method A pp. 231-234.

12) Phosphate - Dissolved Hydrolyzable Phosphorus - FWPCA Official Interim Methods for Chemical Analysis of Surface Waters, September 1968 pp. 230-231.

13) Solids

  a) Dissolved Solids - "Standard Methods" 12th Ed., 1965 Method B pp. 245-246. (Drying Temperature 105°C)

  b) Total Suspended Solids - "Standard Methods" 12th Ed., 1965 Section C pp. 424-425. (Asbestos Fiber used for volatile suspended solids, glass fiber pads used for other tests)

  c) Supernatant Suspended Solids - 1 hour settling (solids measured by preceeding method).

  d) Total Solids - "Standard Methods" 12th Ed., 1965 Method A p. 423.

  e) Volatile Suspended Solids - "Standard Methods" 12th Ed., 1965 Method D p. 425.

14) Sulfates - "Standard Methods" 12th Ed., 1965 Method B pp. 290-291.

15) Total Alkalinity - "Standard Methods" 12th Ed., 1965 pp. 369-370.

16) Total Hardness - "Standard Methods" 12th Ed., 1965 Method B pp. 147-152.

APPENDIX I

Quality Control Program

An analytical quality control program based on the Federal Water Quality Administration's Analytical Quality Control Manual[18] was followed during the latter part of the study. Precision and accuracy control charts were prepared from analytical results obtained by CF&I chemists. Since all but one of the chemists were unfamiliar with wet chemistry methods, a training program was necessary. The quality control program was instituted as part of the training so that each chemist could see if his work compared to actual known values and if he could analyze duplicate samples within the precision limits of the test procedures used.

Precision and accuracy control charts were prepared for phenol, cyanide and ammonia nitrogen since these were the mose important pollutants involved in the study.

Figure 3 is an example of the control charts used during the study. This is a precision control chart for phenol which shows the differences obtained in phenol concentration for duplicate samples. The chart shows the data going out of control on the lower limit indicating increased efficiency. At this point, preparation of a new control chart was necessary using recent data.

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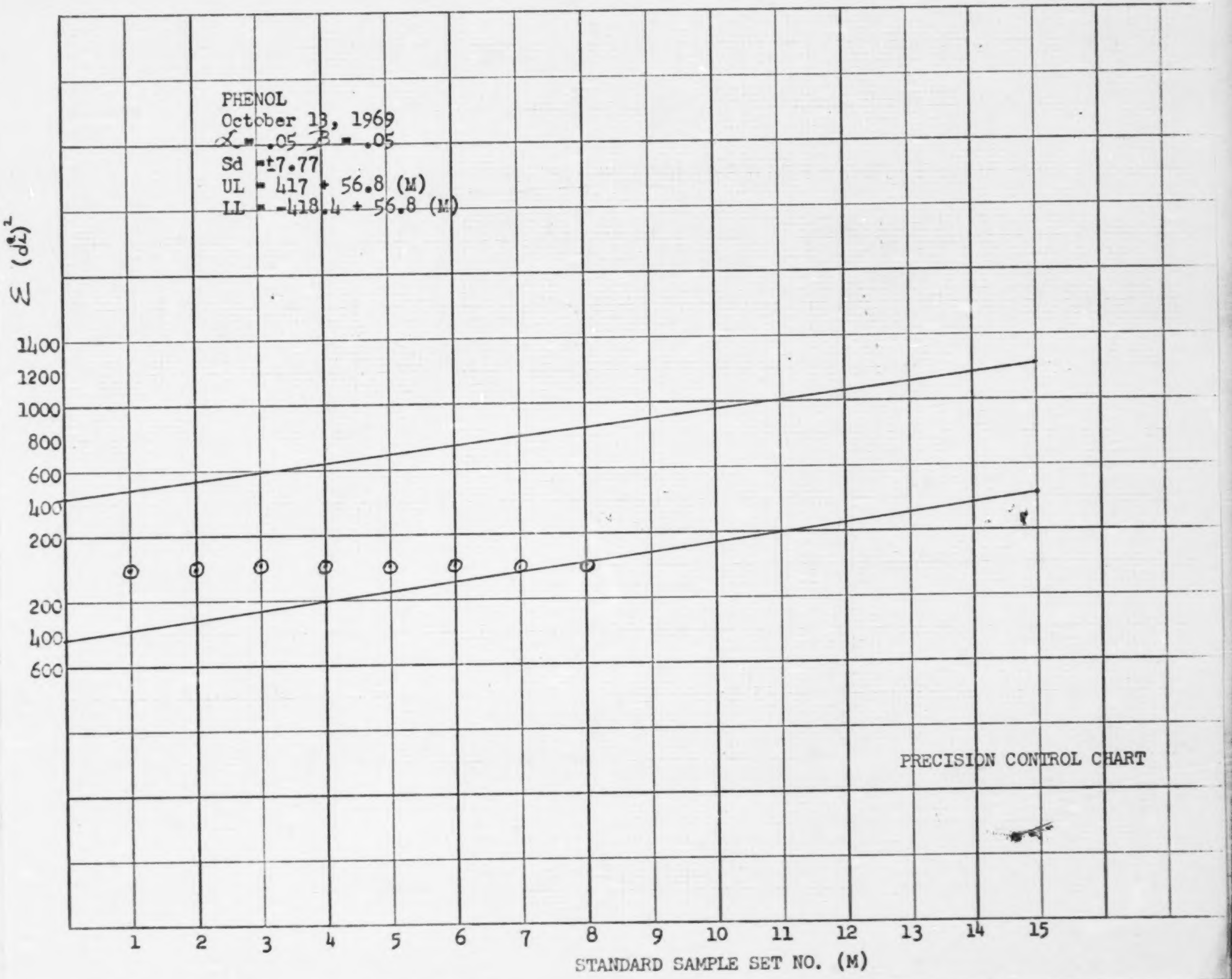


PRECISION CONTROL CHART

APPENDIX J

COAL PREPARATION PLANT DESIGN DATA

[illegible] CORPORATION

JOB RECAP SUMMARY

PUEBLO, COLORADO

ESTIMATE NO.

CUSTOMER [illegible]

JOB B-38650

SHEET 1 OF

DATE 3/31/70

BY C.P.

	WEIGHT OR QUAN.	HOURS	LABOR	MATERIAL	OTHER	TOTAL
JOB RECAP SUMMARY						
ESTIMATED ELECTRICAL						$10 807
ESTIMATED MECHANICAL						382 114
ESTIMATED CIVIL-STRUCTURAL						51 170
ESTIMATED COST TOTAL JOB						$444 091

CORPORATION

ESTIMATE NO. ELECTRICAL SUMMARY SHEET 2 OF

CUSTOMER COLORADO FUEL & IRON PUEBLO, COLORADO DATE 3/31/70

JOB B-38650 BY C.P.

	WEIGHT OR QUAN.	HOURS	LABOR	MATERIAL	OTHER	TOTAL
		ELECTRICAL SUMMARY				
ELECTRICAL BARE COST			$ 2 711	$ 5 096		
TAXES & INSURANCE 14%			380	------		
SUB TOTAL			$ 3 091	$ 5 096		
CONTRACTOR O.H. 20%			618	1 019		
SUB TOTAL			$ 3 709	$ 6 115		
CONTRACTOR PROFIT 10%			371	612		
ESTIMATED TOTAL COST			$ 4 080	$ 6 727		$ 10 807

CORPORATION

ESTIMATE NO.

CUSTOMER COLORADO FUEL & IRON

JOB B-38650

SHEET 3 OF

DATE 3/31/70

BY H.McG. & C.P.

	WEIGHT OR QUAN.	HOURS	LABOR	MATERIAL	OTHER	TOTAL
ELECTRICAL						
150 H.P. MOTOR - INSTALL			300	900		1 200
MOTOR AIR COMPRESSOR STA. & SWITCH			100	310		410
CONVEYOR MOTOR INSTALL SHP.			15	35		50
CENTRIFUGE MOTOR 125 H.P.			300	900		1 200
POLY ELECTROLITE MOTORS (2)			200	210		410
ACID STORAGE PUMP MOTOR 3/4 HORSE			100	100		200
FILTRATE PUMP			125	175		300
275 DENVER CELLS 2 (1/4 H.P.)			50	160		210
LIGHTING (L.S.)			1 020	1 000		2 020
20 H.P. MOTOR - INSTALL			366	958		1 324
10 H.P. MOTOR - INSTALL			135	348		483
			2 711	5 096		7 807

STRUCTURAL SUMMARY

ESTIMATE NO.

CUSTOMER COLORADO FUEL & IRON PUEBLO, COLORADO

JOB B-38650

SHEET 4 OF

DATE 3/31/70

BY C.P.

	WEIGHT OR QUAN.	HOURS	LABOR	MATERIAL	OTHER	TOTAL
			SUMMARY			
STRUCTURAL BARE COST			$ 15 715	$ 20 850		
TAXES & INSURANCE 14%			2 200	-----		
SUB TOTAL			$ 17 915	$ 20,850		
CONTRACTOR O.H. 20%			3 583	4 170		
SUB TOTAL			$ 21 498	$ 25 020		
CONTRACTOR PROFIT 10%			2 150	2 502		
ESTIMATED TOTAL COST			$ 23 648	$ 27 522		$51 170

CORPORATION

ESTIMATE NO.
CUSTOMER COLORADO FUEL & IRON PUEBLO, COLORADO
JOB B-38650
SHEET 5 OF
DATE 3/31/70
BY P.E.L. & C.P.

CIVIL-STRUCTURAL	WEIGHT OR QUAN.	HOURS	LABOR	MATERIAL	OTHER	TOTAL
SUPPORT - BIRD FILTER						
STEEL			6 000	7 800		13 800
STAIRWAY			440	1 760		2 200
NEW WALKWAY CK PL			25	100		125
STRUCTURAL TEAR-OUT			300	----		300
FOUNDATION ACID STORAGE			50	100		150
ANCHOR BOLTS - MONO PUMP			-----	30		30
ANCHOR BOLTS - ELECTROLITE TANKS			-----	150		150
SUPPORT 8 x 8 TANK			200	400		600
CONVEYOR SUPPORT			150	150		300
BASE VACUUM PUMP			80	100		180
COAL FILTER BASE			200	800		1 000
WALKWAY			50	325		375
STAIRWAY			40	200		240
HANDRAIL			40	200		240
AIR COMPRESSOR BASE			100	100		200
SUPPORT FOR TRAP			75	75		150
CONCRETE BASIN	265 C.Y.		6 625	7 950		14 575
CONCRETE SLEEPERS & PUMP PITS	22 C.Y.		400	610		1 010
EXCAVATION	750 C.Y.		940			940
TOTAL BARE COST			$ 15 715	$ 20 850		$35 565

ESTIMATE NO. MECHANICAL SUMMARY SHEET 6 OF
CUSTOMER COLORADO FUEL & IRON PUEBLO, COLORADO DATE
JOB B-38650 BY

	WEIGHT OR QUAN.	HOURS	LABOR	MATERIAL	OTHER	TOTAL
			MECHANICAL SUMMARY			
MECHANICAL BARE COST			$ 59 318	$ 216 232	$ 5 625	
TAXES & INSURANCE 14%			8 305	-----	------	
TOTAL			$ 67 623	$ 216 232	$ 5 625	
CONTRACTOR O.H. 20%			13 525	43 246	1 125	
TOTAL			$ 81 148	$ 259 478	$ 6 750	
CONTRACTOR PROFIT 10%			8 115	25 948	675	
ESTIMATED TOTAL COST			$ 89 263	$ 285 426	$ 7 425	$382 114

278

SHEET 7 OF

ESTIMATE NO.

CUSTOMER COLORADO FUEL & IRON PUEBLO, COLORADO DATE 3/31/70

JOB B-38650 BY C.P.

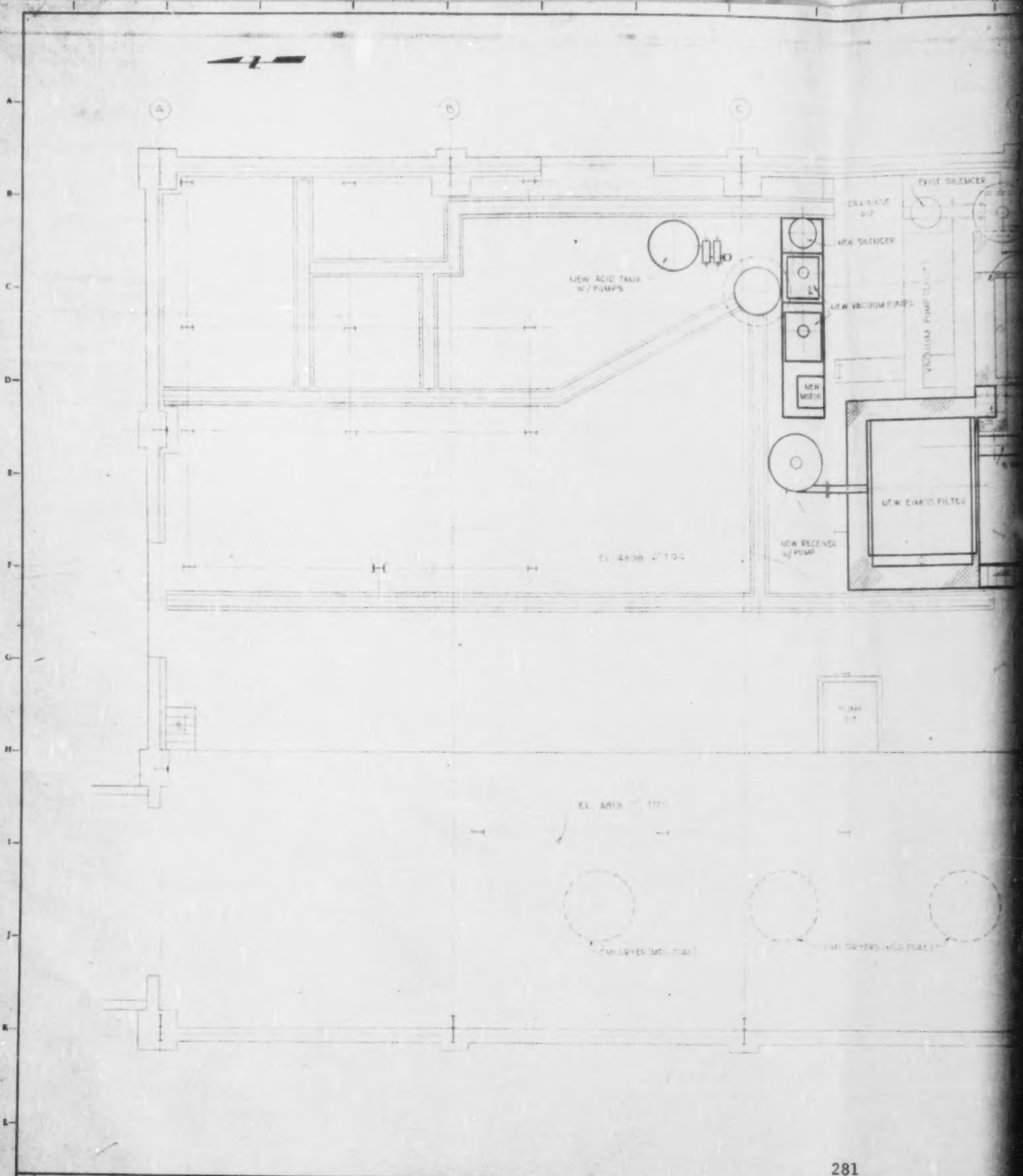
279

	WEIGHT OR QUAN.	HOURS	LABOR	MATERIAL	OTHER	TOTAL
MECHANICAL						
NEW VACUUM PUMP,						
VACUUM TANK,						
FILTER TO PUMP,						
COAL FILTER.			8 640	53 760	1 535	63 935
COAL CONVEYOR			1 800	10 000		11 800
POLY ELECTROLITE TANKS			945	6 750		7 695
AGITATOR			325	2 263	50	2 638
TANK			150	1 442	50	1 642
INTERNALS			375	2 073	100	2 548
DENVER CELLS			1 180	7 356	400	8 936
LEVEL CONTROL			225	1 568	75	1 868
AIR COMPRESSOR			125	1 368	70	1 563
CENTRIFUGE			6 600	47 191	2 600	56 391
MOTOR			510	5 700		6 210
ACID STORAGE			472	3 375		3 847
MONO PUMP			600	6 000	300	6 900
TEAR OUT			4 000			4 000
CHEM. FEED CONTROLS			300	5 000	200	5 500

ESTIMATE NO.
CUSTOMER COLORADO FUEL & IRON PUEBLO, COLORADO
JOB B-38650

SHEET 8 OF
DATE
BY

		WEIGHT OR QUAN.	HOURS	LABOR	MATERIAL	OTHER	TOTAL
	MECHANICAL						
	CRANE			1 100	2 800	200	4 100
	PIPE INSTALLED, HANGERS, SUPPORTS			14 512	11 592		26 104
	FITTINGS & VALVE INSTALLATION						
	VALVES (MATERIAL ONLY)				8 796		8 796
	PIPING FOR FROTHING AGENTS			900	1 100		2 000
	PIPING-VACUUM SYSTEM			1 848	1 462		3 310
	TEAR OUT			3 000			3 000
280	REWORK CHEM. FROTHING FEEDS			1 500			1 500
	PUMP WITH MOTOR			100	957	45	1 102
	STEEL SHELL & DISCHARGE CONE			5 474	14 494		19 968
	MECHANICAL & BRIDGE			4 637	21 185		25 822
	TOTAL BARE COST			59 318	216 232	5 625	281 175


A
B
C
D
E
F
G
H
I
J
K
L
NEW ACID TANK W/PUMPS
DRAINAGE PIT
EXIST SILENCER
NEW SILENCER
NEW VACUUM PUMPS
VACUUM PUMP
NEW MOTOR
NEW EIMCO FILTER
NEW RECEIVER W/PUMP
EL 4808 TOC
PUMP PIT
EL 4815 TOC
EMI DRYER (NO. ...)
EMI DRYERS (NO. ...)

A

NO	REVISIONS	DATE	BY	CK'D	APP'D	REFERENCE DRAWINGS

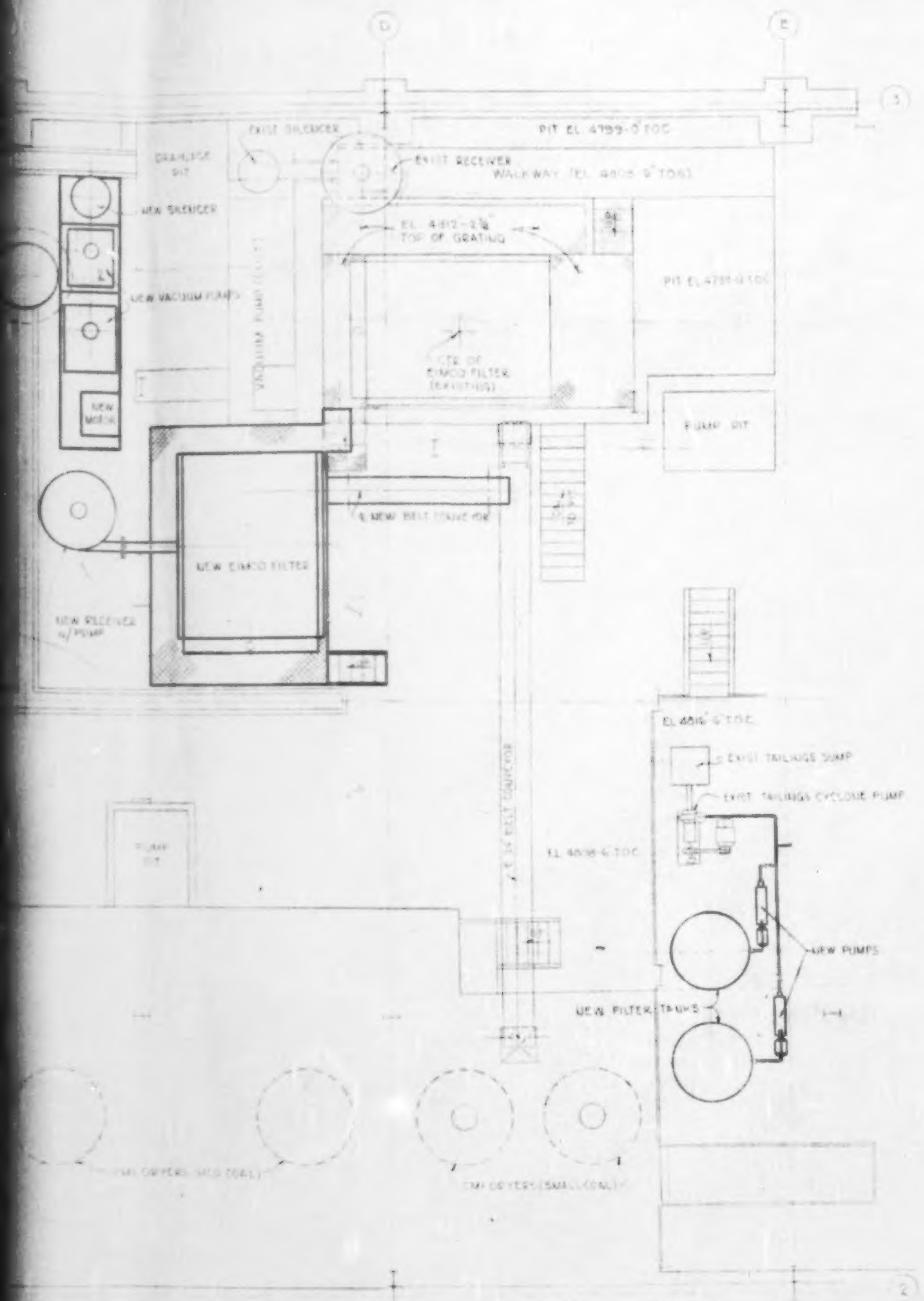
PRINT RECORD

REVISION NO

ISSUED

CUSTOMER

FIELD


D
E
3
EXIST SILENCER
DRAINAGE PIT
EXIST RECEIVER
PIT EL 4799-0 T.O.C.
WALKWAY (EL 4808-9" T.O.S.)
NEW SILENCER
EL 4812-2⅞" TOP OF GRATING
NEW VACUUM PUMPS
VACUUM PUMP (EXIST)
PIT EL 4799-0 T.O.C.
CTR OF EIMCO FILTER (EXISTING)
NEW MOTOR
PUMP PIT
℄ NEW BELT CONVEYOR
NEW EIMCO FILTER
NEW RECEIVER w/PUMP
EL 4816'-6" T.O.C.
EXIST TAILINGS SUMP
EXIST TAILINGS CYCLONE PUMP
PUMP PIT
℄ E 34 BELT CONVEYOR
EL 4808-6 T.O.C.
NEW PUMPS
NEW FILTER TANKS
CMI DRYERS (MED COAL)
CMI DRYERS (SMALL COAL)
2

B

PRINT RECORD

Stearns-Roger CORPORATION

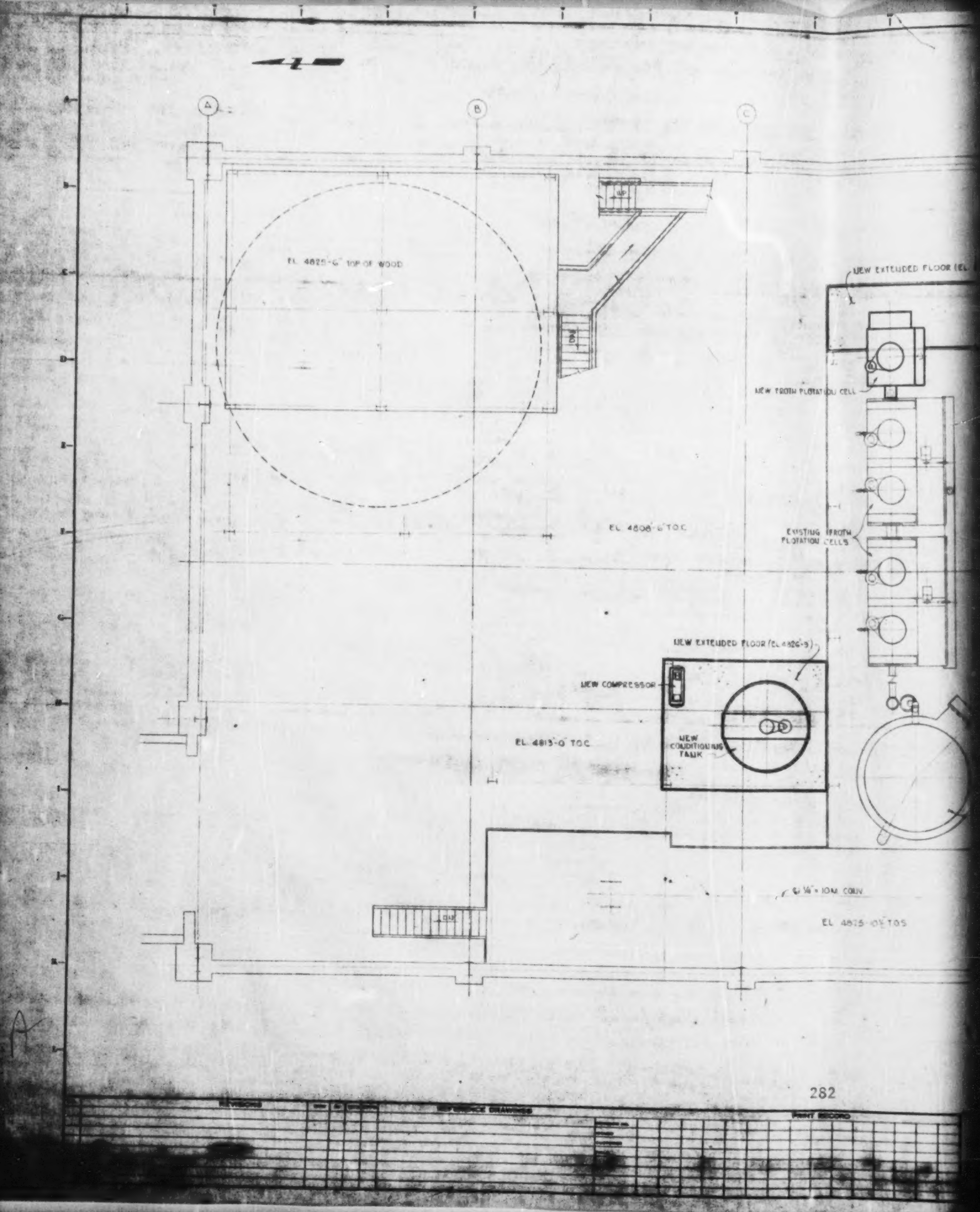
C.F. & I. CORP. PUEBLO, COLO.

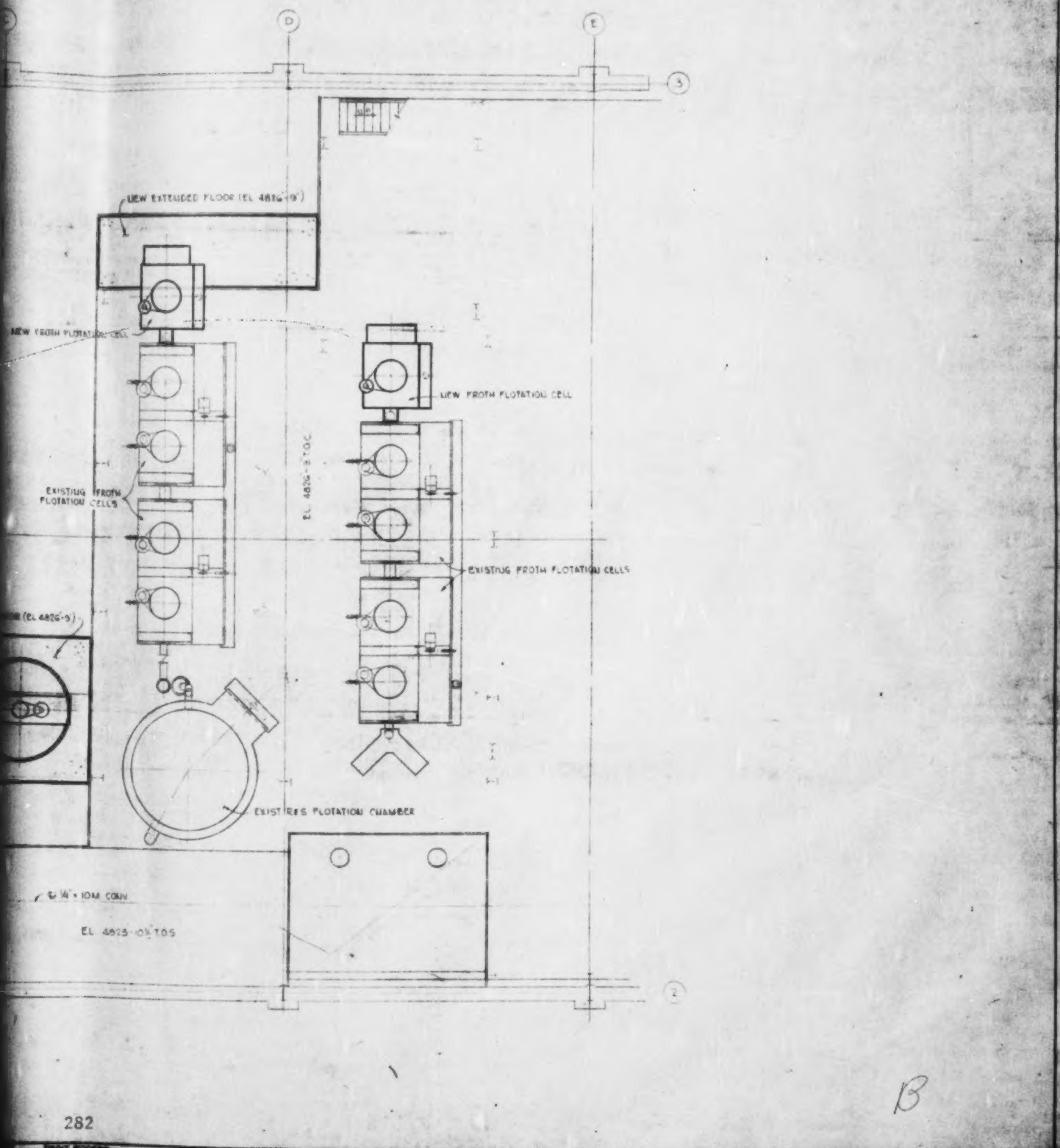
[redacted]

TITLE
NEW POLLUTION CONTROL EQUIPMENT

PLAN AT ELEVATIONS
4808'-6" & 4816'-6"

4


A
B
C
UP
DN
EL. 4825'-6" TOP OF WOOD
EL. 4808'-6" T.O.C.
EL. 4813'-0" T.O.C.
EL. 4825'-10" T.O.S.
NEW EXTENDED FLOOR (EL.
NEW FROTH FLOTATION CELL
EXISTING FROTH FLOTATION CELLS
NEW EXTENDED FLOOR (EL. 4826'-9")
NEW COMPRESSOR
NEW CONDITIONING TANK
℄ 1/4" = 10M CONV.
UP
REVISIONS
REFERENCE DRAWINGS
PRINT RECORD


D
E
3
UP
NEW EXTENDED FLOOR (EL 4826'-9")
NEW FROTH FLOTATION CELL
NEW FROTH FLOTATION CELL
EL 4826'-5" T.O.C.
EXISTING FROTH FLOTATION CELLS
EXISTING FROTH FLOTATION CELLS
(EL 4826'-9")
EXISTING FLOTATION CHAMBER
¼" × 10M CONV
EL 4825'-0½" T.O.S.
2

B

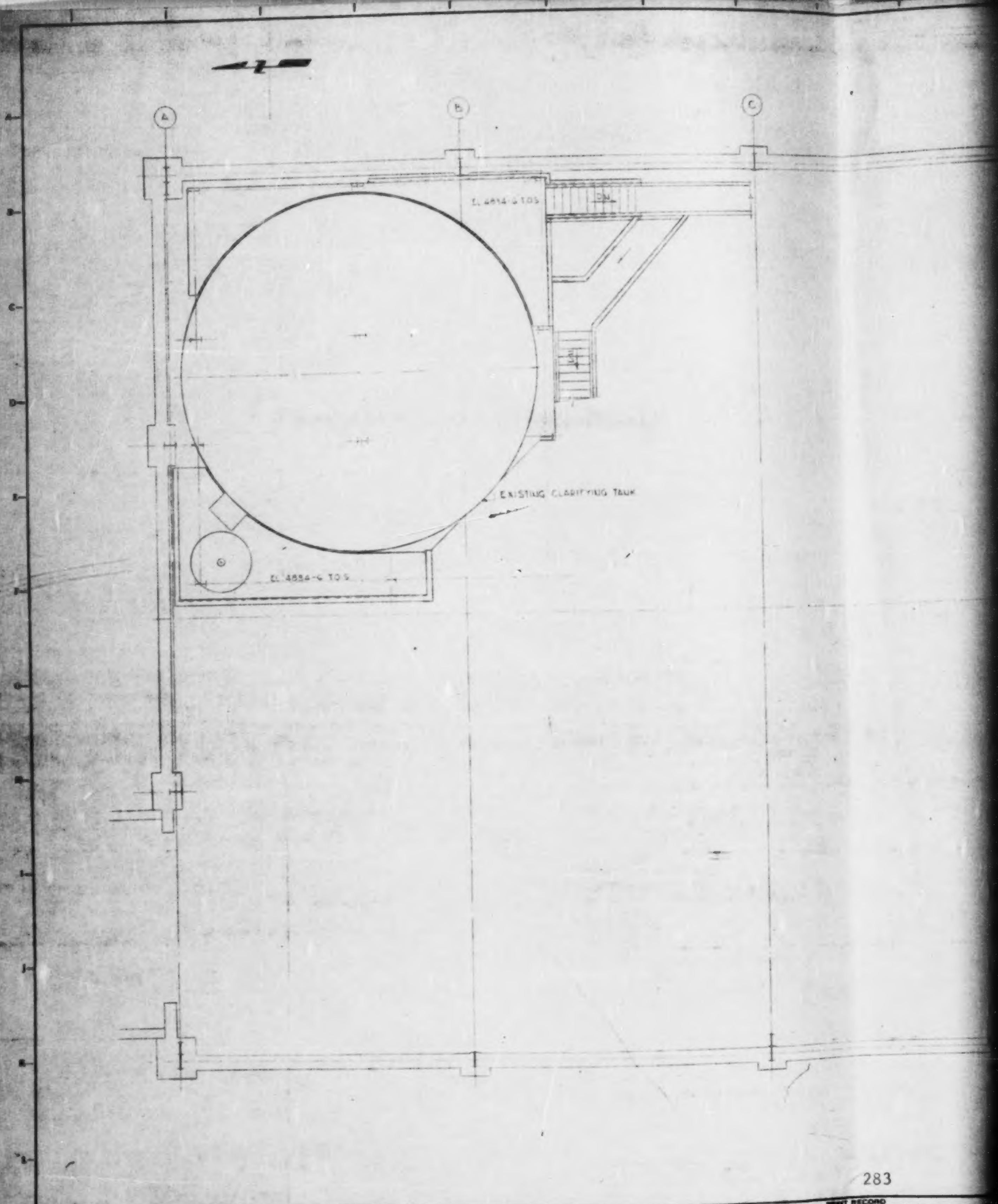
PRINT RECORD

Stearns-Roger

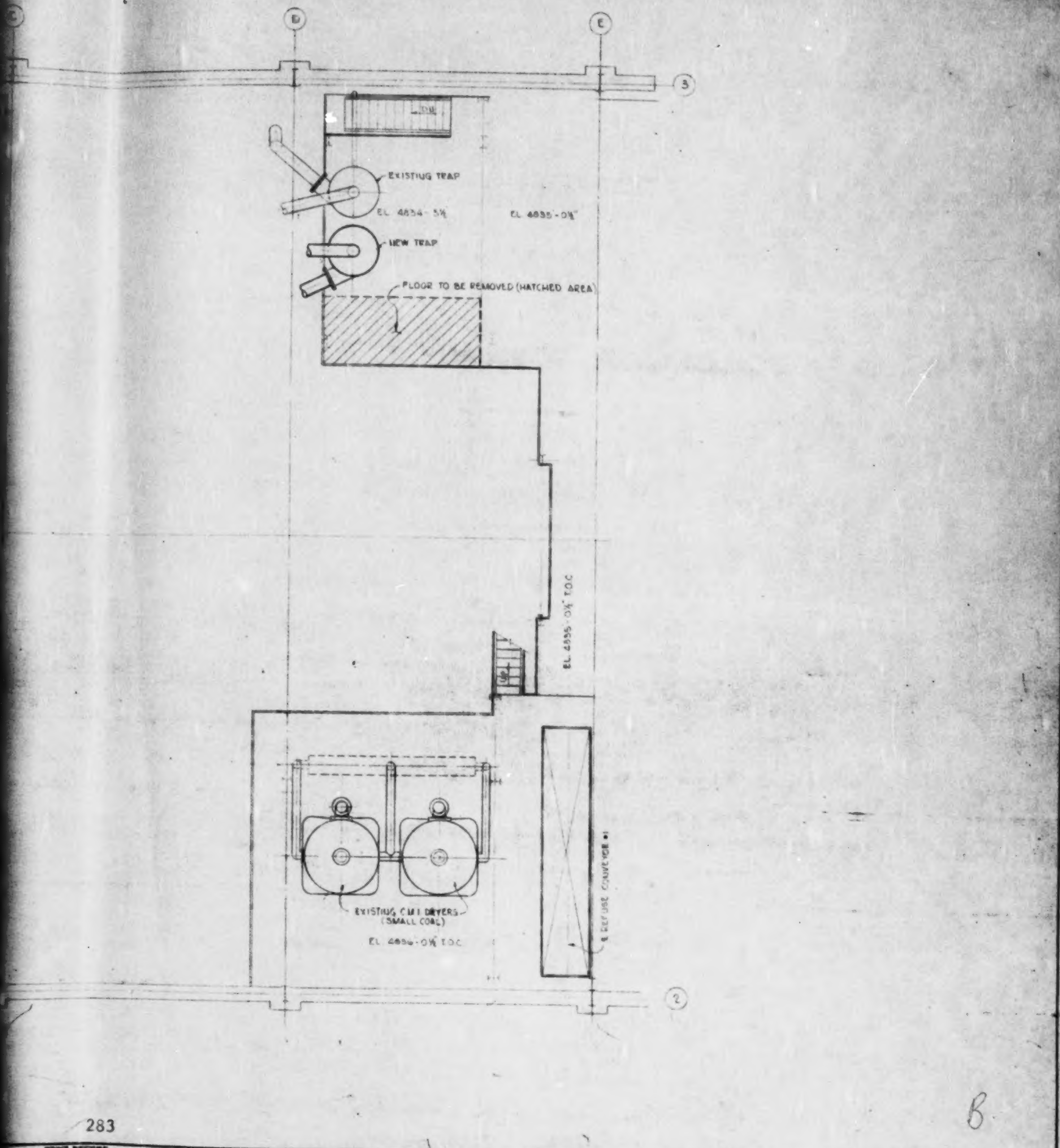
C.F. & I. CORP. PUEBLO, COLO

TITLE
NEW POLLUTION CONTROL EQUIPMENT

PLAN AT ELEVATION
[illegible]


A
B
C
EL 4854-6 T.O.S
DN
UP
EXISTING CLARIFYING TANK
EL 4854-6 T.O.S
A
B
C
D
E
F
G
H
I
J
K
L
A

NO.	REVISIONS	DATE	BY	CKD	APPD	REFERENCE DRAWINGS	PRINT RECORD
							REVISION NO.
							ORDER
							CUSTOMER
							FIELD


C
D
E
3
2
DN
EXISTING TRAP
EL 4834-3½
EL 4835'-0½"
NEW TRAP
FLOOR TO BE REMOVED (HATCHED AREA)
EL 4835'-0½" T.O.C.
UP
EXISTING C.M.I. DRYERS (SMALL COAL)
EL 4836-0½ T.O.C.
℄ REFUSE CONVEYOR #1

283

B

PRINT RECORD

Stearns-Roger

C.F. & I. CORP. PUEBLO, COLO

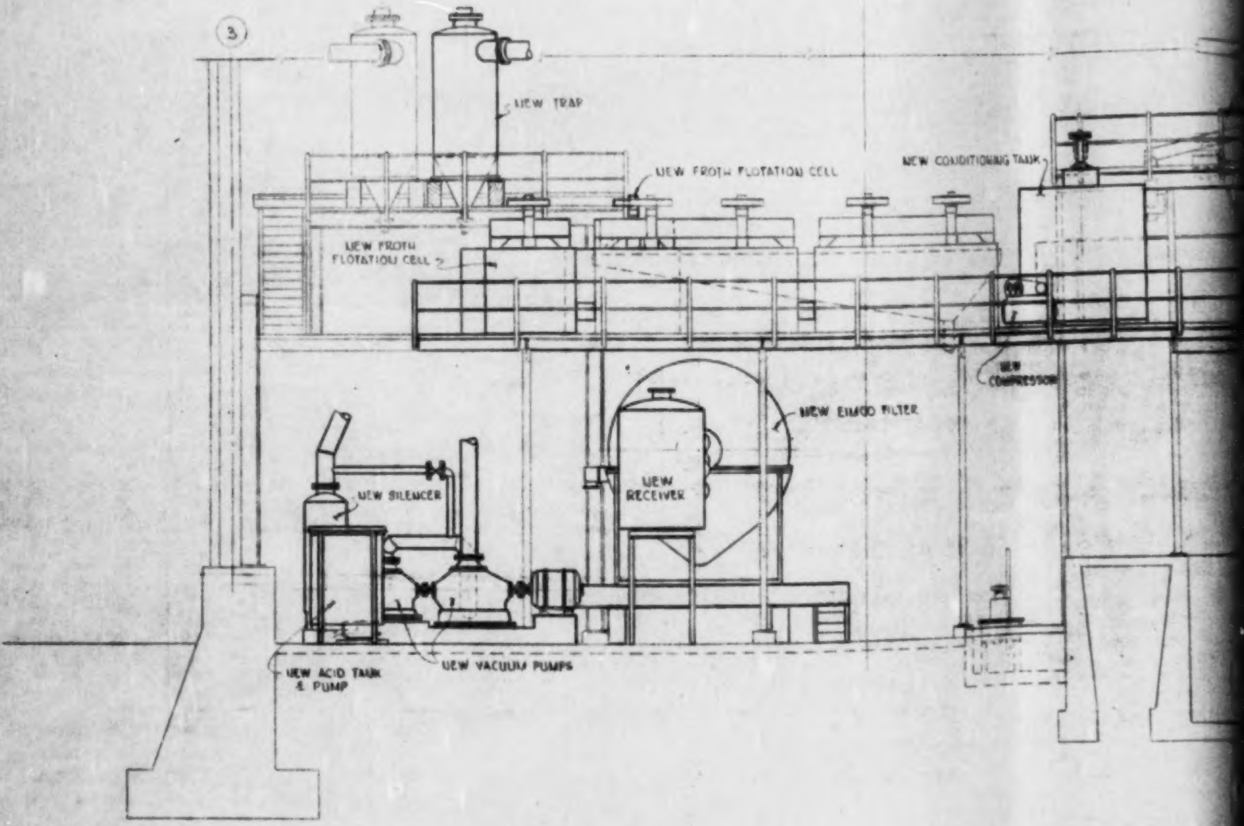
J SAVAGE 3-27-[illegible]

TITLE
NEW POLLUTION CONTROL EQUIPMENT

PLAN AT ELEVATIONS
4834'-3½" & 4836'-0½"

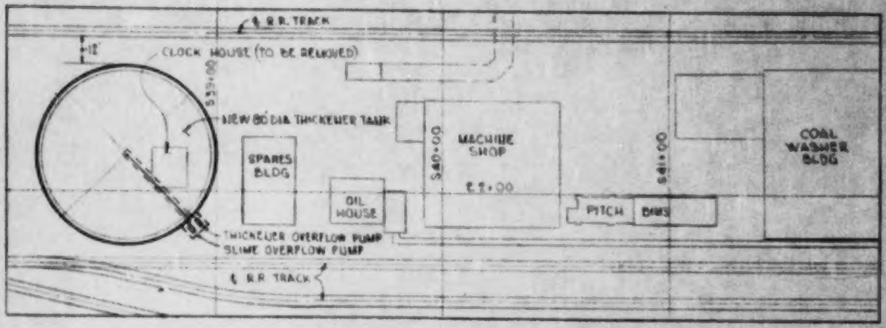
SCALE ¼"=1'-0"

3


3
NEW TRAP
NEW FROTH FLOTATION CELL
NEW FROTH FLOTATION CELL
NEW CONDITIONING TANK
NEW COMPRESSOR
NEW EIMCO FILTER
NEW SILENCER
NEW RECEIVER
NEW ACID TANK & PUMP
NEW VACUUM PUMPS

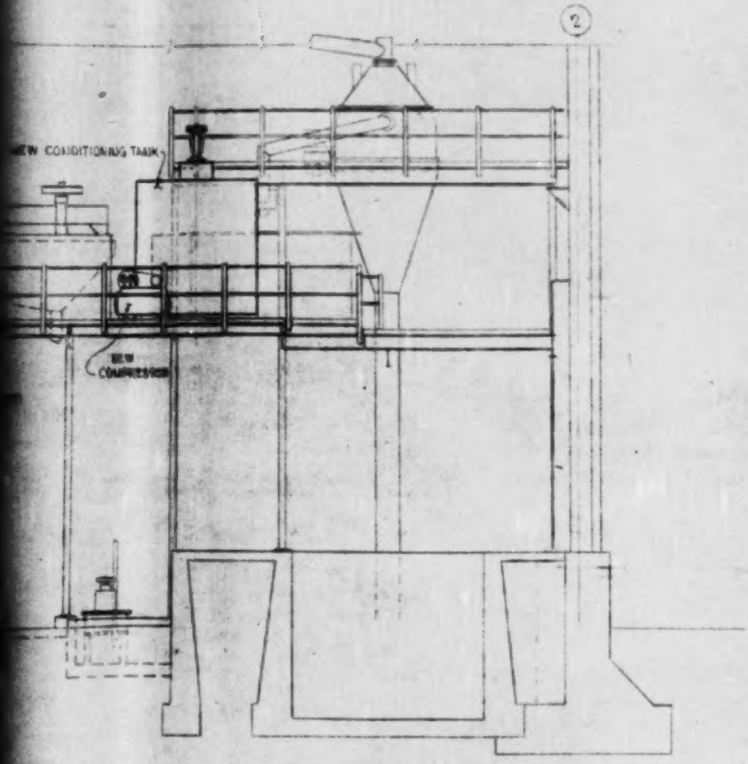
SECTIONAL ELEVATION VIC. COL. LINE C - LOOKING SOUTH
SCALE: ¼" = 1'-0"

REVISIONS	DATE	BY	CHK'D	APP'D	REFERENCE DRAWINGS	PRINT RECORD
						REVISION NO.
						ISSUED
						CUSTOMER
						FIELD



℄ R.R. TRACK
12'
CLOCK HOUSE (TO BE REMOVED)
539+00
NEW 80' DIA THICKENER TANK
SPARES BLDG
OIL HOUSE
MACHINE SHOP
540+00
E 2+00
541+00
PITCH BINS
COAL WASHER BLDG
THICKENER OVERFLOW PUMP
SLIME OVERFLOW PUMP
℄ R.R. TRACK

LOCATION PLAN - THICKENER TANK
SCALE: 1" = 30'



2
EW CONDITIONING TANK
NEW COMPRESSOR

INE C - LOOKING SOUTH

B

284

PRINT RECORD

Stearns-Roger

C.F. & I. CORP., PUEBLO, COLO.

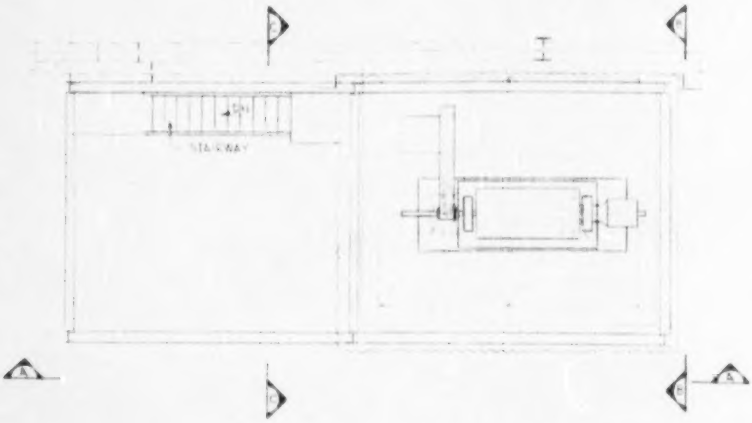
J. SAVAGE 4-1-70

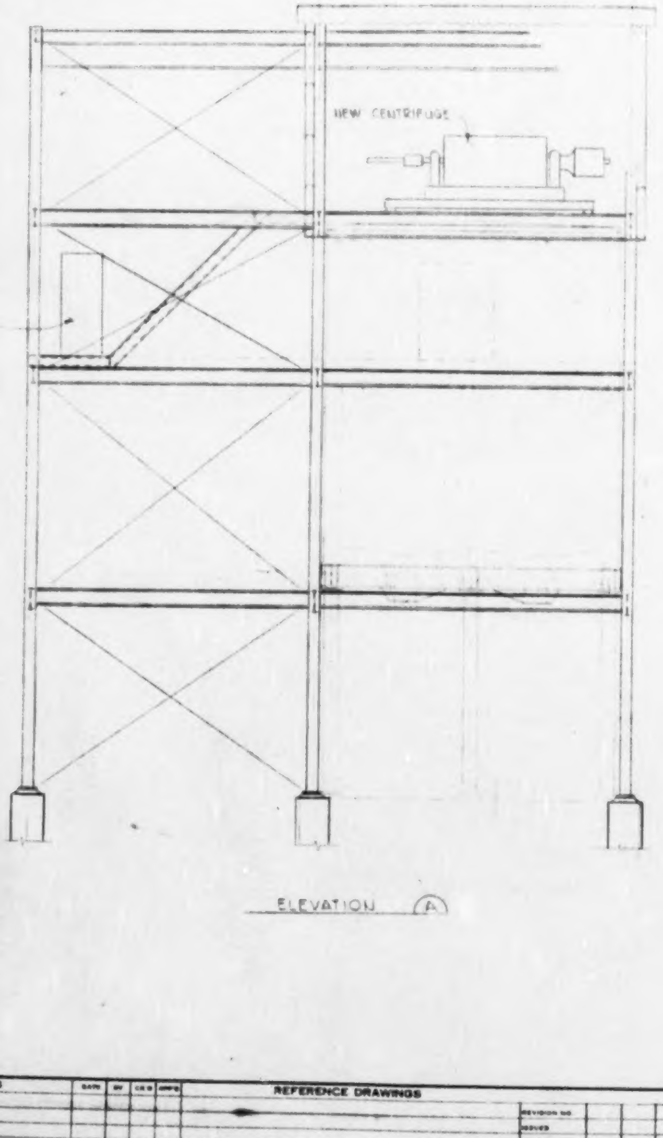
TITLE
NEW POLLUTION CONTROL EQUIPMENT.

SECTIONAL ELEVATION,
INTERIOR COMPONENTS
& PLAN OF EXTERIOR THICKENER

AS NOTED

4

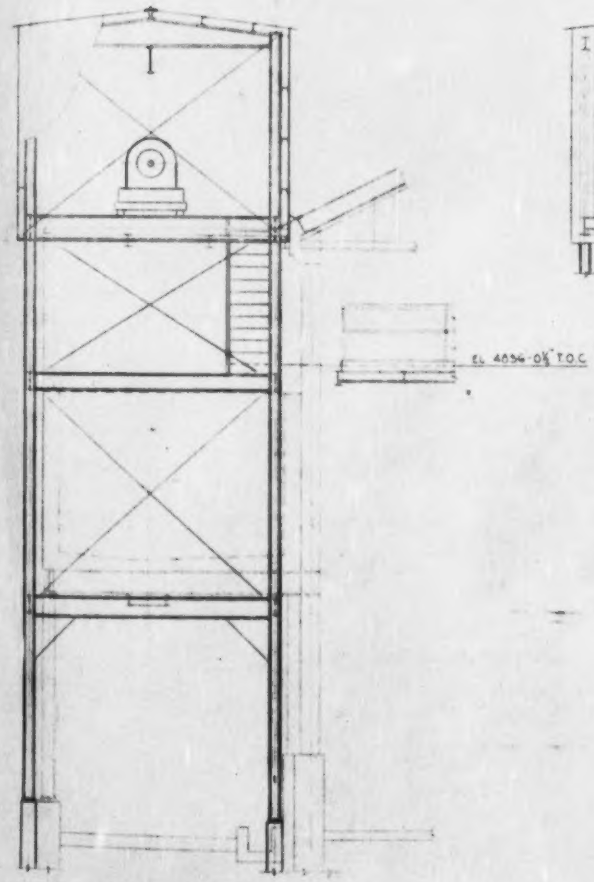

B
B
DN
STAIRWAY
A
A
B
B


NEW CENTRIFUGE
NEW DOOR IN
EXISTING WALL

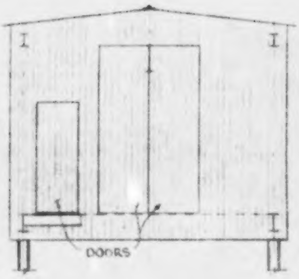
ELEVATION (A)

A

NO.	REVISIONS	DATE	BY	CKD	APPD	REFERENCE DRAWINGS	PRINT RECORD
							REVISION NO
							ISSUED
							CUSTOMER
							FIELD



ELEVATION (B)



SECTION (C)

B

PRINT RECORD

Stearns-Roger

C.F. & I. CORP. PUEBLO, COLO

DRAWN J. SAVAGE 4-9-70

TITLE
NEW POLLUTION CONTROL EQUIPMENT

PLANS & ELEVATIONS
FOR CENTRIFUGE ADDITIONS

SCALE ⅛" = 1'-0"

5 △

APPENDIX K

Ion Exchange Resin Data

Amberlite 200 and Amberlite IRC-84 Resin Data

	Amberlite 200 Lot 4069	Amberlite IRC-84 11-D-515
Total Exchange Cap.		
meq/ml	1.74	4.13
meq/ml	5.00	10.00
Salt Splitting Cap.		
meq/ml	--	0.25
meq/ml	--	0.62
Water Retention Capacity %	55.1	49.8
Wet Screen Analysis		
+16	1.8	1.7
+20	15.1	19.1
+30	43.6	52.1
+35	22.6	11.8
+40	14.5	9.7
+50	2.1	5.2
+100	0.3	0.4
Effective Size mm	0.46	0.48
Pressure Factor	511	493
Swelling H+ Na %	--	59.0
Na H+ %	4.0	--